HOT GAS CLEANUP OF GASIFICATION PRODUCTS

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SUMMARY

A study has been made of particle cleanup from gases produced in an atmospheric pressure 0.305-m dia. coal gasifier. Three devices were explored - a conventional cyclone, a vortex pocket cyclone of the Cardiff design and a granular bed filter of coarse particles operated in upflow.

Typical solids loadings leaving the gasifier reached 0.3 kg solids/kg gas (300,000ppmw) when solids were recirculated to the bed from the primary cyclone. These solids were separated in a hot primary cyclone at inlet velocities of about 17 m/s to give an exit stream of about 2500ppmw. Particles recovered from the primary cyclone were about 65% carbon with the balance ash. Mean particle size collected was about 35μm for coal tested, and 65μm for oil sands cokes.

The vortex pocket cyclone was operated at a very high inlet velocity of about 120 m/s and lower inlet loadings of 30,000 to 100,000ppmw with which outlet concentrations averaged 360ppmw over nine experiments.

Eight tests were done with a granular bed filter downstream of the vortex pocket cyclone. The bed was about 0.25-m deep and packed with collector particles 2.2mm in diameter. Inlet loadings were about 350 to 650ppmw. Efficiencies were scattered and low, ranging from 15 to 57%. These low values are in the range calculated from literature models for the given conditions with particles below 2μm in size. Particles collected were composed largely of ash, and contained about 15% carbon. Further work is clearly indicated for the granular bed as collection efficiencies are critically dependent on particle size in the range of interest.

Scanning electron microscopy was found to be useful in interpretation of size distribution data for the fine particulates being measured by different techniques. Photomicrographs of solid samples collected in the different configurations tested, and passing through the collection system are presented.

Results in terms of inlet and outlet solids loadings were in reasonable agreement with the few studies available in the literature for coal gasification systems.

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1. INTRODUCTION

Combined cycle power generation based on coal gasification requires that gases be cleaned of particulates before combustion to minimize gas turbine erosion problems. If the gases can be cleaned while hot, increased cycle efficiencies can be achieved compared to situations where the gas is cleaned at ambient temperatures. Hot gas cleanup also has advantages for cycles based on pressurized fluidized bed combustion. Thus there is a growing interest in cleaning gases of particulates arising from coal in both combustion and gasification processes. Papers at two recent conferences, "Fluidized Bed Combustion - Is It Achieving It's Promise", Institute of Energy, London, 1984 and "Gas Cleaning at High Temperatures", Inst. Chem. Eng., University of Surrey, Surrey, England, Sept. 1986, focus on the hot gas cleanup problem and provide a state-of-the-art assessment of current technology.

Turbine specification are reported by Dellefield and Bedick[1] as requiring particulate loadings below 0.02 g/nm^3 *for particles smaller than 5μm, and no particles larger than 5μm. Rubow[2] reports pressurized fluidized bed combustion turbine tolerance limits as follows:

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* For atmospheric pressure gasifiers, product gas density at 15°C is about 0.9-1.1 kg/m^3, thus solids loadings expressed as g/nm^3 when multiplied by 1000 give approximate values in ppmw.

Turbine Manufacturer	Internal Limits	Projected Total
General Electric	1ppmw > 12.7μm 4ppmw > 8μm 8ppmw > 5μm	200ppmw
Westinghouse		40-100ppmw
Stal-Laval	12ppmw > 5μm	200ppmw

It has been argued that for combined cycles based on gasification, particulate levels in the clean gas could be considerably higher than for cycles based on pressurized fluid bed combustion since the particles are carbonaceous in nature and will be partially consumed in the combustion of the gas prior to the flue gas being sent to the turbine. For combined cycles based on coal combustion the flue gas goes directly to the turbines and the above specifications should be met.

Hot gas cleanup usually involves primary collection of particulates by one or two stages of cyclones, followed by a secondary cleanup device which will hopefully approach the turbine specification limits. These secondary collectors may be advanced cyclone designs, augmented by electrostatic devices, granular beds, electrostatic precipitators, ceramic filters and the like.

1.1 Conventional Cyclones

Given that primary collection is usually by cyclone, some features of cyclone separation should be reviewed. The major limitation of cyclones is their low efficiency for collecting particles smaller than 5μm. Even if the cyclone is efficient at room temperature, as has been shown[3,4], collection efficiency will drop as temperature increases. Further, as the barrel diameter of a cyclone increases above 0.3m the collection efficiency decreases. Flocculation can occur in cyclones where loadings are high ($230 g/nm^3$) resulting in high efficiency of removal of particle sizes in the 0.1 to 2μm range. Cyclone efficiency generally increases as the dust load to the cyclone increases. Inlet loadings up to several kg/m^3 have been reported. Collection efficiencies of properly designed cyclones approach 100% for particles larger than 50μm in diameter, but reach values of below 50% for particles of the order of a few microns in size.

Cyclone efficiencies can be predicted for different conditions by equations such as those of Leith[5].

$$\eta = 1 - \exp\left[-2(C\psi)^{\frac{1}{2n+2}}\right]$$

where $$\Psi = d_p^2 \rho_p V(n+1)/18\mu D$$

$$n = 1-(1-0.67D^{0.18})(T/283)^{0.3}$$

and C is a constant governed by the geometrical ratios of the cyclone, and is given in Appendix B.

A typical efficiency plot for primary and secondary cyclones operating in a high temperature combustion environment reported by Wheldon et al.[6] is shown as Figure 1. While the primary cyclone is well over 95% efficient, the secondary cyclone shows much lower efficiencies.

1.2 Solids Loadings from Gasifiers

For coal gasifier outlet streams particulate levels have been reported as below. Oakley and Reid[7] report as typical the following particulate properties after cyclone separation from a fluidized bed gasifier as

Solids loading	2000ppmw
Solids composition	70% wt. Carbon, 30% wt. Ash
Particle size	d_{50} 6μm

Dellefield and Bedick[1], in studying secondary cleanup devices, report on the 0.76-M dia. General Electric pressurized stirred bed gasifier. A 0.254-M dia. cyclone was used with inlet loadings of 5400ppmw at 7 bar, and outlet loadings of 104ppmw at 580°C. After the one cyclone, the particles analyzed 90% wt. Carbon, and the remainder presumably ash. A bench scale 0.152-M dia. fluid bed gasifier was used by the same authors along with a particulate feeder to increase loadings for testing secondary devices. They state as typical, loadings of 14.1g/nm^3 (14,000ppmw) and 0.7 g/nm^3 (700ppmw) in fluid bed gasifiers with one and two cyclones respectively. They tested secondary devices with inlet concentration of 265 to 7975ppmw.

Hozumi et al.[8] report results with a 40 ton/day fluid bed gasifier. After two stages of hot cyclones and a hot desulphurizer they report inlet dust levels to granular bed filters as $3g/nm^3$ (3000ppmw). For a fluid bed combustor Makhejee et al.[9] quote levels of 5000-22000ppmw for flue gas loadings and 200ppmw after three stages of cyclones.

This survey suggests cyclone outlet loadings below 14,000ppmw for fluidized bed gasifiers after a single cyclone, and in the range 3000ppmw and below for two stages of cyclones. For a moving bed gasifier such as the G.E. stirred bed design after one cyclone a value of 100ppmw is expected. If one assumes an efficiency of 95% for these cyclones inlet concentrations to the primary cyclones of a fluid bed gasifier will be about 280,000ppmw or 0.28kg solids/kg gas.

Expected composition of solid char particles from a fluid bed gasifier is between 70% C (Oakley and Reid) and 90% C (Dellefield and Bedick).

2. Objectives of the Research Program

The objectives for the current work were to explore features of hot gas cleanup from a pilot scale atmospheric pressure coal gasifier. Testwork involved three devices (i) standard cyclones (ii) a novel vortex pocket cyclone based on the Cardiff-design and (iii) a granular bed. Different configurations of these units were to be evaluated, with operation at varying velocities, and in the case of the granular bed, with different particle size, composition and bed depths. Compositional and size distribution data for the collected particles as well as the material escaping the collection devices was to be obtained. Gasifier feedstocks of coal and coke were to be used.

The specific tasks for the contract are given in Appendix A.

3. Experimental Apparatus and Procedures

3.1 The Gasifier

Hot gas cleanup was explored using the U.B.C. atmospheric pressure 0.305-M inside diameter spouted bed gasifier (Figure 2). This unit has been described in detail elsewhere[10], however, a brief description is given below.

Table 1 lists some critical dimensions of the reactor system. The reactor itself consists of three major sections; the main body is a 10 gage carbon steel column with outside diameter 0.50 m and height 1.37 m, and is refractory lined to an inside diameter of 0.30 m to withstand the high temperatures. Below the main section is a 0.46 m high 316 stainless steel cone tapered at 50 degrees which is also refractory lined. The uppermost section of the reactor is a larger diameter (0.76 m) disengaging section, which reduces the amount of particulates leaving the gasifier.

During the pre-heating stages of the experiment, hot gas enters the cone bottom and flow upwards spouting and heating a pre-determined charge of bed material. This flue gas is generated in the propane burner, as shown in Figure 2.

TABLE 1: Gasifier Specifications

Gasifier:	
Column	0.50-m O.D. x 0.305-m I.D. x 1.37-m high
Cone	Angle 60°, height 0.46-m
Disengaging Section	0.76-m O.D. x 0.56-m I.D. x 0.61-m high
Spouted Bed Orifice Diameter	3.2 cm
Startup Burner	176 kW max. propane fired
Cyclone - 1st Stage:	20.6 cm dia. x 27.3 cm cylindrical height x 33.7 cm cone height
Cyclone - 2nd Stage:	19.4 cm dia. x 26.7 cm cylindrical height x 48.3 cm cone height
Gas Coolers (3)	Each 2.2 m length x 7.6 cm I.D. x 10 cm dia. jacket
Gas Knock Out Drum	56 cm dia. x 90 cm height

Once the reactor has been heated to the desired temperature, coal or coke is pneumatically transported from the vibra-screw feeder into the reactor via the 2.54-cm dia. solids transport line. It is also possible to feed a second solid material (either dolomite or potassium carbonate) into the solids transport line by means of the auxiliary feeder/rotary valve assembly.

During gasification, flue gas and some solids carry over from the reactor into the gas cleanup section. In the usual operation with conventional cyclones particulate material is removed by the two cyclones in series, and when necessary, dust can be recycled from the first cyclone into the reactor through a rotary valve. Char from both cyclones is collected in a char (dust) drum and weighed and analyzed following each experiment. Hot gas leaving the cyclones is cooled in an indirect heat exchange with water and any remaining dust is removed in a knockout drum and filter. The cleaned gas can then be sampled and analyzed before the excess is sent to a propane-fired incinerator located on the roof of the building.

3.2 Gasifier Operating Procedure

Before the start of the gasification experiment the gas chromatograph is checked to ensure it is operating properly and if necessary it is recalibrated. The various bottled gases, including propane, pure oxygen and quenching nitrogen, are then connected to their respective piping. Air from the building air compressor is fed to the inlet of the reactor, the coal feed line and the char recycle line to prevent them from being blocked with bed material. Depending on the experimental conditions, the bed material may

consist of char from a previous run, silica sand or dolomite. The initial static bed depth of a typical run is about 0.45 m and the bed material is charged from the top of the reactor. The reactor is then heated to 500°C using the propane fired preheater. The heating time varies from 1 hour for the char bed to 2 hours for the silica sand or dolomite bed. The bed is spouted using air before the coal is blown into the reactor using pure (undiluted) oxygen. As the reactor temperature increases the propane to the propane preheater is decreased gradually and finally shut off when the bed temperature reaches 750°C. The spouting air is gradually replaced with saturated steam. At 0.5 hr after the introduction of coal, if desired, the char recycle system is turned on and the char is conveyed using air to the reactor. The reactor temperature is then controlled at the desired value by adjusting the steam and oxygen flow rates.

One hour after the start of the experiment, gas samples are taken at 20 minute intervals and analyzed for H_2, CO_2, O_2, N_2, CH_4, CO, H_2S and SO_2 using gas chromatographs. Sampling of gas for solids loading begins when the reactor temperature is steady, which is about 2 hrs into the run. The temperatures, pressures, and various flowrates of the gasifier system are also recorded at the same times as the gas is sampled for chromatographic analysis. A typical experiment lasts form three to five hours to ensure that a steady state gas composition is achieved. The coal feed, pure oxygen, char recycle feed, air and steam are shut off at the desired time. To prevent any fire hazard the ball valves for the coke feed and char recycle lines are turned off (see Figure 2). The reactor is purged with a short burst of nitrogen to remove any trace of oxygen from the reactor. All electrical

power to the gasifier is turned off and nitrogen, pure oxygen and propane are shut off at their respective sources. The reactor is allowed to cool overnight.

The next day, the char collected by the cyclones is discharged from the bins and weighed. A sample is analyzed for carbon and ash content. The coal feeding system is cleaned and the weight of coal used during the experiment is determined. The dust filter is dismantled and cleaned for the next run. The reactor cone is lowered and the bed material is weighed. The carbon and ash content for the bed sample is also determined by a combustion test. Then the reactor cone is place back into position for the next run. If the pressure drop during the experiment is excessive, (>124 kPa) then the piping in the system is checked and any blockages are cleaned in preparation for the next run. The data collected are processed using the computer program and the results are analyzed. Particle size distributions of the solids found in the bed, and the cyclone catch are determined by sieving. In certain cases the sludge from the dust filter is analyzed for ash, and carbon and its size distribution determined.

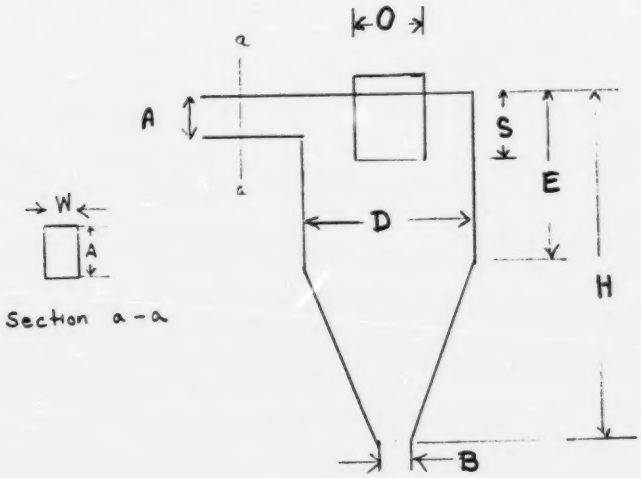
3.3 The Conventional Cyclones

Prior to the start of this contract, the gasifier was equipped with two cyclones in series as shown in Figure 2. Details of the cyclones are given in Figure 3, and their critical ratios listed in Table 3. In comparison with "standard" cyclones of various recommended designs the conventional cyclones are seen (Table 3) to resemble a Swift high efficiency cyclone except that the cone of cyclone No. 1 is foreshortened somewhat to give an H/D ratio of about 3 rather than 4. Cyclone No. 2 is essentially a

TABLE 3: Comparison of Standard Cyclones with Recommended Designs*

Ratio	Stairmand High Efficiency	Swift General Purpose	Swift High Efficiency	U.B.C. No.1	Cyclone No.2
A/D	0.50	0.50	0.44	0.59	0.43
W/D	0.20	0.25	0.21	0.21	0.21
O/D	0.50	0.50	0.40	0.38	0.38
S/D	0.50	0.60	0.50	0.59	0.5
E/D	1.50	1.75	1.40	1.33	1.38
H/D	4.0	3.8	3.9	3.0	3.9
B/D	0.375	0.40	0.40	0.37	0.39

* From Fayed and Otten[11], Stairmand[12], Swift[13]



Swift design. These cyclones were both available prior to starting this contract and were used with no further modifications except for the addition of sampling probes to be discussed below.

Material collected by the cyclone was determined by weighing the quantities collected in the cyclone receivers. These were typically of the order of 20 kg or more for the primary cyclone.

3.4 The Vortex Pocket Cyclone

As part of this work, a vortex pocket cyclone was fabricated. Dimensions are given in Figure 4. The design is based on the Cardiff cyclone being studied by the U.K. Coal Research Establishment and described by Biffen et al.[14], Sage et al.[15], and more recently by Syred et al.[16]. This design is claimed to be 94% efficient for particles of 5.7μm, 76% for 3.6μm dia. particles and 55% for 2.3μm dia. particles. The inlet cross-sectional area of this cyclone is reduced over that of the conventional cyclones used, which increases considerably the inlet velocity. The object of evaluating the vortex cyclone was to see whether enhanced efficiencies in collection of small particles could in fact be achieved under gasification conditions.

Material collected by the cyclone was determined by weighing that found in the cyclone receiver. This stream was of the order of 20kg for a typical experiment.

3.5 Granular Bed Filter

The granular bed collector is shown in Figure 5. Granular bed filters can be designed in many different configurations, but most are variants of two basic approaches - the cross-flow design, or the counter current flow design. Since the collector particles must ultimately be cleaned of dust, they must move through the collector, usually by gravity. The gas flow may be across a shallow bed held in place by louvers, such as is shown in Figure 6a, (Shaw et al. 17). In this configuration the face velocity is usually low, and the bed shallow in the direction of flow. In the counter current flow configuration with down flow of collector particles, a conical bottom section is required to permit exit of the collector particles. Beds are usually deeper in the flow direction than the cross flow case. A typical example is shown by Guillorg[18] (Figure 6b). The design selected for the current study was a variant of the counter current flow design, because of its inherent simplicity of construction and operation given the space available at the UBC gasifier site. Geometric factors are listed for the filter in Figure 5.

The amount of solids collected on the particles was determined by cleaning the collector particles as described below in section 5.

3.6 Sampling of Gas Streams for Solids Loading

The particular configuration of collection devices and the locations of the sampling probes used to isolate the performance of each will be described under their respective sections of the report. Sample probes were designed to extract samples of the flowing solids laden gas. Figure 7 shows

the basic construction. The probe itself has a cross-sectional area of 0.1642×10^{-4} m^2 as compared to the pipe in which it is inserted which has a flow area of 30.87 $10^{-4} m^2$. It is constructed of a 6.35mm O.D. dia. tube of 316 stainless steel, bent at a 90 degree angle to the flow and cut off at right angles to the flow. It was placed at the centre line of the flow.

The sampling train is based on configurations outlined by Cadle[23] modified as necessary for the current conditions. For most runs the gas was extracted through a high temperature ceramic filter, through a moisture knock-out drum, a diaphragm pump, and then is exhausted through a wet test meter. The choice of the wet test meter rather than a dry flowmeter was made because of the very high moisture contents of the gases being sampled when the oxygen-steam blast was used in gasification. The flow of sample gas was then measured under saturated conditions at room temperature. It was not possible to verify that the flowrate was completely constant over the sampling periods which ranged from ten to sixty minutes although the wet test meter dial rotation was kept under observation in case of blockage of the ceramic filter.

Solids loadings were calculated as g/m^3 of saturated gas at room temperature, and then calculated back to actual loadings at temperature knowing the total gas mass flowrate.

During operation it was impossible to maintain isokinetic conditions because of unknown moisture content and gas yield, and the necessity to modify steam rates to control temperature. However, as will be discussed in section 5.2, well established corrections for isokinetic sampling conditions

could be applied to correct the results as necessary. The magnitude of the correction is a function of the particle size being collected among other variables.

Appendix C gives a sample calculation for the solids loading determination.

3.7 Particle Size Determinations

Particle size distributions were measured uisng a Particle Data Inc. Elzone (Coulter-type) counter Model 112LSRD /SSP/ADC-80XY/TP 50. Particles are dispersed in an electrolyte made up of 0.45μm distilled water, 0.5% wt. tetrasodium pyrophosphate, and 0.75% wt. sodium chloride. For most of the work a 93μm orifice with a useful particle diameter range of 1.4-1.9μm to 38-47.5μm was used. If necessary the 380μm orifice with a useful particle diameter range of 5.7-7.6μm to 150-190μm was used. The coulter counter technique gives an equivalent sphere diameter based on volume.

For coarser particles such as primary cyclone products, sieving was used as a check on the coulter counter results. Size distribution data was also determined using a light microscope with motorized stage image analyzer. This unit has associated software which gives mean values and variances. The diameters determined here are those of a projected sphere and the distributions are by number, not by volume or mass. Thus values are not in a strict sense comparable to Coulter counter measurements. As will be discussed in section 5, scanning electron microscopy was used to interpret size distribution results by the above three methods.

3.8 Particle Composition

Composition of particles is important as it influences particle density, which affects the performance of most collection devices. In this work since the particles in question are chars, they were analyzed in most cases only for % ash, with the remainder being assumed to be carbon. Analysis was done by combustion in a muffle furnace. For some cases where quantities were insufficient (e.g. dust collected on granular bed collector particles) carbon concentrations were determined commercially by microanalytical techniques.

3.9 Gasifier Feedstocks

Analysis of the coals and cokes used in the experiments are given in Table 4. Particle size distribution of feed and typical spouted bed materials are given in Table 5. Testwork was done using oil sands coke, a northern B.C. anthracite, and an Alberta sub-bituminous coal.

4. Scope of the Test Program

The overall scheme of the hot gas cleanup studies is shown in Table 6. Runs listed are those specifically done for this contract alone, plus runs from contract OST/84-00055 and contract OST85-00217 where extra work was done to analyze the loadings to the cleanup devices, and the size distribution of collected solids. Runs 134 to 142 were used to characterize the base case operation where the original two essentially standard design high efficiency cyclones were used in series. Runs 143 to 149 used the standard high

TABLE 4: Feedstocks Used in Hot Gas Cleanup Studies

Feedstock	Mt. Klappan Anthracite-1	Mt. Klappan** Anthracite-2	Highvale Sub-bit. Coal	Suncor Coke	Syncrude Coke
lemental Analysis % Wt. Dry Basis)					
C	60.94	66.22	58.68	84.9	79.49
H	2.01	2.06	3.87	3.93	1.56
N	0.79	1.06	0.88	1.32	1.69
S	0.40	0.33	0.15	6.02	6.96
Cl	0.06	0.01	0.02	0.01	0.70
O*	3.38	3.42	15.76	0.73	1.30
Ash	32.42	26.89	20.64	3.09	8.30
roximate Analysis % Wt. As Rec'd)					
Moisture	7.16	5.48	19.79	1.80	3.73
Ash	30.11	25.42	16.56	3.03	7.99
Volatiles	7.69	6.99	28.63	11.94	6.89
Fixed Carbon	55.06	62.11	35.02	83.23	81.39

** Used for Runs 158 - 160

* by difference

TABLE 5: Size Distributions of Gasifier Feeds

Feedstock	Mt. Klappan Anthracite-1		Highvale Sub-bit. Coal		Suncor Coke		Syncrude Coke	
	d_i (mm)	x_i	d_i (mm)	x_i	d_i (mm)	x_i	d_i (mm)	x_i
	4.76	.007	+4.76	.0106	+4.76	0.09		
	4.05	.0638	3.56	.2104	3.56	0.49		
	2.85	.1502	2.03	.1685	2.03	0.116		
	2.03	.1765	1.44	.1588	1.44	0.112		
	1.35	.2600	1.02	.1260	1.02	0.091		
	0.80	.165	0.780	.0512	0.72	0.045	0.71	.0297
	0.478	.0816	0.568	.0977	0.6	0.146	0.475	.0591
	0.303	.0346	0.363	.0567			0.302	.1744
	0.215	.0337	0.216	.0367			0.214	.5157
	0.153	.0181	0.181	.0329			0.163	.0909
	0.116	.0036	0.128	.0169			0.106	.1204
	0.106	.0059	0.0795	.0247			0.063	.0098
			0.0455	.0062				
			0.0380	.0027				
$d_p = \frac{1}{\Sigma \frac{x_i}{d_i}}$ (mm)	0.831		1.748		1.49		0.20	

x_i = Weight Fraction

d_i = avg. screen aperture size (mm)

TABLE 6: Scheme of Hot Gas Cleanup Studies

Run	Feedstock	Gasification Blast	Hot Gas Cleanup Device		Recycle from Primary
			Primary	Secondary	
134	Syncrude coke	O_2 - Steam	Std. Cyclone-1	Std. Cyclone-2	No
135	"	"	"	"	Yes
136	"	"	"	"	"
137	Suncore Coke	"	"	"	"
138	"	"	"	"	"
139	Mt. Klappan Anth.	"	"	"	No
142	"	O_2 - Steam	Std. Cyclone-1	Std. Cyclone-2	Yes
143	"	O_2 - Steam	Std. Cyclone-1	Vortex Pocket Cy	Yes
144	"	"	"	"	No
145	"	"	"	"	Yes
146	"	"	"	"	"
147	"	"	"	"	"
149	"	"	Std. Cyclone-1	Vortex Pocket Cy	Yes
150	"	O_2 - Steam	Vortex Pocket Cyc	Std. Cyclone-2	No
151	"	Air - Steam	"	"	"
152	Mr. Klappan Anth.	"	"	"	"
153	Highvale Coal	"	"	Granular Bed	"
154	"	"	"	"	"
155	"	"	"	"	"
156	"	"	"	"	"
157	Highvale Coal	Air - Steam	"	"	"
158	Mt. Klappan Anth.	O_2 - Steam	"	"	"
159	"	"	"	"	"
160	Mt. Klappan Anth.	O_2 - Steam	Vortex Pocket Cyc	Granular Bed	No

efficiency cyclone as the primary collector and the vortex pocket (Cardiff cyclone) unit as the secondary. Runs 150 to 160 used the vortex pocket cyclone as the primary collection device. In runs 150 to 152 the standard high efficiency cyclone was the secondary device, while in Runs 153 to 160 the granular bed filter was used. Recycle from the primary collector was used in about half of the runs. This results in approximately a doubling of the inlet loading to the primary cyclone. Recycle of solids was not used in any runs with the vortex pocket cyclone as primary collector.

From the viewpoint of feedstocks, five runs are reported with coke as feed, fourteen with an anthracite whose high ash content assures considerable particulates, and five runs with a sub-bituminous coal.

5. Results and Discussion

5.1 Gasification Results

Gasification conditions, product gas analyses and gasification performance indicators are summarized in Tables 7 and 8. The tests covered a wide range of conditions in terms of reactor temperatures, feed rate, bed material, blast condition, velocity and feedstocks. While the gasification results per se are not the primary focus of this report, a few comments are in order.

For runs with recycle of solids from the primary cyclone, even where the blast is oxygen and steam, the air assisting the recycle results in a dry gas containing about 15% N_2 by volume, with a corresponding decrease in dry gas heating value.

TABLE 7 Gasification Results for Highvale Coal and Oil Sand Coke

RUN NO	FEED STOCK	BED CONTENT (INITIAL)	BED WEIGHT (kg)		FEEDRATE (kg/h)					m_{O_2}/m_C	m_S/m_C	U_O (m/s)	TEMP (°C)		GHV MJ/m³	GY m³/kg	% C CONVERSION	DRY GAS ANALYSIS (%v)				
			INITIAL	FINAL	m_{O_2}	m_C^*	m_{N_2}	m_S	m_{CH}				AVG	MAX				H_2	CO	CO_2	CH_4	N_2
153	HV	CHAR	16.4	7.2	21.3	12.2	70.2	36.4	0.0	1.75	2.99	1.79	983	1031	0.62	5.59	74.8	2.2	1.8	17.6	0.04	78.3
154	HV	"	13.4	9.8	31.4	25.1	103.2	19.8	0.0	1.25	0.79	2.63	958	1006	2.75	5.19	92.4	9.6	11.2	14.2	0.63	64.4
155	HV	"	16.4	7.7	39.5	28.0	129.9	23.9	0.0	1.41	0.86	3.08	986	1032	2.14	5.56	81.9	7.1	9.4	15.6	0.39	67.5
156	HV	"	16.4	8.9	20.8	19.0	72.0	5.9	0.0	1.15	0.31	1.74	984	1035	2.80	4.88	89.8	8.3	12.9	13.6	0.5	64.7
157	HV	CHAR	16.5	8.6	22.0	17.4	72.3	6.7	0.0	1.26	0.38	1.40	911	974	2.62	4.11	76.4	7.9	12.5	14.2	0.4	65.0
134	SY	DOL	43.6	18.1	25.4	25.9	15.3	47.9	0.0	0.98	1.85	1.54	1036	1059	1.51	1.50	57.8	9.1	3.7	67.4	0.0	20.2
135	SY	DOL	38.6	46.4	23.5	29.4	9.4	39.3	29.8	0.80	1.34	1.36	940	950	4.42	1.74	55.5	23.2	12.8	44.3	0.33	19.3
136	SY	DOL	40.8	28.7	23.8	22.4	9.1	38.4	29.8	1.06	1.72	1.44	858	872	6.50	2.83	80.5	33.1	20.4	32.5	0.06	13.9
137	SU	CHAR	18.5	27.7	24.4	21.8	8.2	36.3	16.3	1.12	1.67	1.52	914	925	5.89	3.33	96.2	29.2	15.1	37.3	1.39	17.0
138	SU	CHAR	27.6	54.4	23.5	20.7	7.9	39.6	16.3	1.13	1.91	1.56	943	947	6.55	2.78	81.0	32.4	19.7	32.4	0.43	15.4

HV = HIGHVALE SY = SYNCRUDE COKE (FLUID BED) SU = SUNCOR COKE (SPOUTED BED) DOL = DOLOMITE RUNS 134-136 HAVE EXTRA DOL. FEED.

* MASF BASIS

TABLE 8 Gasification Results for Mt. Klappan Anthracite

RUN NO.	BED WT. (kg) INITIAL	BED WT. (kg) FINAL	FEED RATE (kg/h) m_{O_2}	FEED RATE (kg/h) m_C^*	FEED RATE (kg/h) m_{H_2}	FEED RATE (kg/h) m_S	FEED RATE (kg/h) m_{CH}	m_{O_2}/m_C	m_S/m_C	U_0 m/s	TEMP (°C) AVG	TEMP (°C) MAX	G.H.V. MJ/m³	G.V. m³/kg	%C CONVERSION	DRY GAS ANALYSIS (%v) H_2	CO	CO_2	CH_4	N_2
139	28.4	74.7	21.1	24.0	0.0	38.6	0.0	0.88	1.61	1.50	888	900	7.63	1.21	45.6	31.8	27.3	39.2	1.27	0.39
140	30.0	53.5	20.0	19.1	0.0	38.1	0.0	1.05	1.99	1.49	913	925	7.76	1.57	58.5	32.3	28.9	36.9	0.87	0.72
141	27.3	30.7	22.6	17.2	7.3	39.7	16.6	1.32	2.32	1.49	972	990	5.40	2.36	76.6	20.8	22.5	40.5	0.46	15.5
142	27.4	34.4	23.7	17.0	11.6	44.4	16.3	1.39	2.60	1.44	932	939	4.43	2.12	70.5	17.4	19.2	38.4	0.34	25.7
143	28.4	31.2	26.1	12.1	12.0	43.4	16.3	2.15	3.58	1.71	1008	1010	4.61	2.52	79.5	20.2	19.0	36.8	0.32	23.7
144	35.9	45.7	22.4	13.0	4.5	52.4	0.0	1.72	4.04	1.66	986	1001	4.20	1.84	69.7	18.3	15.1	51.8	0.52	14.3
145	28.4	30.6	25.6	16.4	11.8	49.7	16.3	1.56	2.72	1.64	1020	1097	5.13	2.39	77.2	23.8	23.2	34.6	0.41	18.0
146	28.3	29.4	25.5	12.0	11.8	40.5	16.3	2.13	3.38	1.54	994	1026	4.16	2.52	82.2	18.4	18.3	39.7	0.35	23.4
147	30.4	36.9	24.4	16.2	12.0	43.0	16.3	1.50	2.65	1.51	906	925	4.59	1.75	56.0	17.3	19.2	37.6	0.46	25.4
149	14.1	32.4	21.4	13.4	5.7	30.9	16.3	1.60	2.31	1.20	980	995	6.20	1.82	63.5	24.1	25.8	36.3	0.44	13.3
150	27.1	42.7	22.3	20.3	7.34	6.8	0.0	1.10	0.33	1.37	970	972	2.55	3.12	47.0	8.3	12.2	15.0	0.20	64.3
151	29.3	19.5	48.8	15.1	16.7	6.6	0.0	3.23	0.44	2.66	859	874	0.0	7.36	66.1	0.0	0.0	10.8	0.0	79.1
152	19.5	30.5	31.0	22.8	102.0	35.3	0.0	1.36	1.54	2.75	964	971	0.86	5.04	63.2	4.5	5.0	17.4	0.14	72.9
158	30.0	66.9	25.7	17.2	12.9	44.6	0.0	1.49	2.58	2.03	968	1013	5.60	3.63	75.6	30.5	21.8	31.1	0.41	16.2
159	21.3	26.5	21.6	19.9	71.2	9.1	0.0	1.49	0.41	1.88	943	1048	2.0	4.88	70.0	7.2	9.4	15.5	0.2	67.9
160	27.1	34.4	23.6	14.5	77.0	27.7	0.0	1.49	1.91	1.35	814	960	2.2	4.25	76.5	9.0	9.5	16.8	0.0	64.6

* MAF BASIS

Dry gas compositions for oxygen-steam gasification without recycle are typically $\%H_2$=32, %CO=28, $\%CO_2$=38, $\%CH_4$=1.3, $\%N_2$=0.7. Carbon conversion, defined as mol($CO+CO_2+CH_4$) in off-gas/mol C fed, tend to increase with temperature. For the anthracite, temperatures in excess of 1000°C are needed for reasonable percentage conversion (Runs 143,145,146), whereas for the sub-bituminous coal carbon conversion was high at lower temperatures (Runs 153-157).

All runs gave reasonable off-gas composition except for conditions where high velocity was first attempted (Run 151) in which feed problems gave air/coal ratios more typical of combustion than gasification, and for the initial run with Highvale coal (Run 153). Wet gas compositions are given in Appendix C.

In some cases problems in the gasifier operation precluded satisfactory sampling of the gases for solids loading. Generally it was very difficult for two operators to satisfactorily monitor the gasifier operation with its needs for gas composition sampling and immediate analysis, as well as give undivided attention to the particulate sampling train. Results in terms of the hot gas cleanup studies will be discussed by configuration, once the gas loading sampling results and collection efficiency calculations are presented.

5.2 Cyclone Efficiency and Gas Loading Calculations

Cyclone efficiency can be calculated from the mass flows of solids according to equation (1).

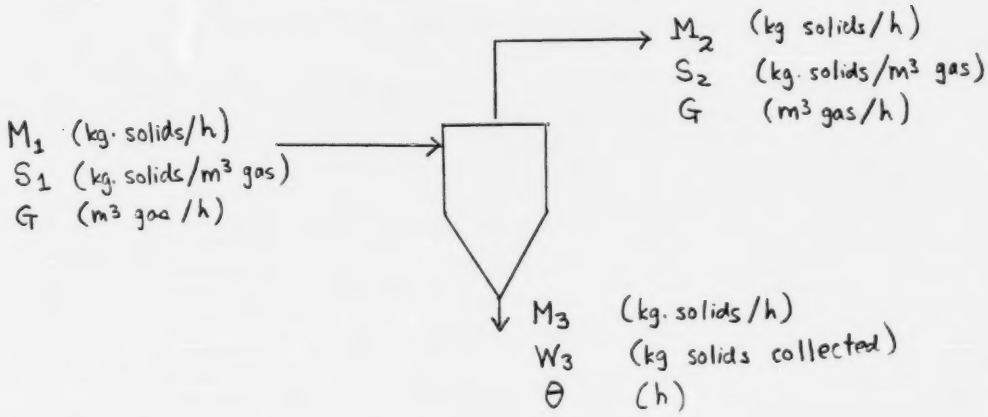
$$\eta = M_3/M_1 \qquad (1)$$

where M = mass flow of solids

Now $M_1 = M_2 + M_3$, so that one could evaluate efficiency as

$$\eta = M_3/(M_2 + M_3) = \frac{1}{1 + M_2/M_3} \tag{2}$$

or $$\eta = (M_1 - M_2)/M_1 = 1 - M_2/M_1 \tag{3}$$



In general M_3 is the most accurate mass flow measurement, being calculated form the large mass of collected solids W_3, over a long time Θ i.e. $M_3 = W_3/\Theta$, whereas M_1 and M_2 are calculated from gas sampling measurements to give the respective solids concentrations S_1 and S_2 and knowing the volumetric flow of gas, G (i.e. $M_1 = S_1G$, $M_2 = S_2G$). Where the cyclone is the primary unit, where the inlet gas loading is high, sampling for S_1 rapidly fills the filter crucible, and blockage may occur, and the resulting values are subject to some error. Thus it seemed preferable to determine the cyclone efficiency from equation (2), rewritten as

$$\eta = \frac{1}{1 + S_2G/M_3} \tag{4}$$

Here, for a 95% efficient cyclone, a ±50% error in S_2 changes efficiency only by about ±2.5%, although the difference in downstream concentration between a 92.5% and a 97.5% efficient cyclone is not insignificant. Where the numbers warranted, efficiency were determined as well from equations (1) and (3), and an average taken.

Values of the solids loading in the gas were determined by sampling the suspension through the probe described in section 3.6. Because it was not possible to know the bulk velocity a priori or to maintain isokinetic sampling conditions (where the sampling velocity was equal to the bulk velocity) corrections for anisokinetic sampling were calculated.

Badzioch[19] established corrections to dust loading values for anisokinetic sampling. For flow through a pipe or stack, if sampling is done at a velocity below the bulk velocity, the calculated dust loading will be affected since some particles that should enter the sampling nozzle go past it rather than enter. By contrast if the sampling velocity is higher than the bulk velocity the quantity of particles entering the nozzle will be higher than characteristic of the flow. However, the error associated with anisokinetic sampling is a strong function of particle inertia. Since small particles follow deflected gas streamlines, the concentration of particles will be equal to those in the gas stream irrespective of sampling velocity. For very coarse particles which are not deflected by flow around the probe, and persist in their original direction of motion, the mass flowrate of particles is measured correctly irrespective of sampling velocity. Thus the magnitude of the correction for anisokinetic sampling is a function of particle inertia.

The equation developed by Badzioch[19] and widely used by others is:

$$W_s/W = (1-\alpha)V_s/V + \alpha \qquad (5)$$

where W_s = mass of particle entering sample nozzle at sampling velocity V_s

W = mass of particles which would enter at velocity V, the bulk velocity in the duct.

This equation is valid over the range $0.5 < V_s/V < 4$ which is within the limits of the present tests (Tables 9 and 14). The parameter α depends on the inertia of the particles. For "small" particles $\alpha \to 0$ and $W_s/W \simeq V_s/V$, and for "large" particles $\alpha \to 1$ and $W_s \simeq W$. The value of α is given by

$$\alpha = [1 - \exp(-L/\lambda)]/[L/\lambda] \qquad (6)$$

where $\lambda = V_t V/g$ with V_t being the terminal velocity and g, the acceleration due to gravity. L, the distance upstream over which the probe disturbs the flow streamlines is a linear function of probe diameter given by Badzioch which for the present probe has a value of 0.052 m.

For the present case, calculations show (Appendix B) that α approaches zero for $d_p < 10\mu m$, and hence $W/W_s \simeq V/V_s$ by equation (5). Thus the mass of particles entering will be proportional to the sampling velocity and the concentration (or loading) in g/nm^3 will be correct. Thus, although the sampling velocities in the current work exceed the bulk velocity in the pipe, the particles are sufficiently small (or have sufficiently low inertia) that the correct value of the concentration should be obtained. Thus no corrections have been used in the body of this report to the measured concentrations of particles in the gas streams. Nevertheless corrected loadings are listed in Appendix B.

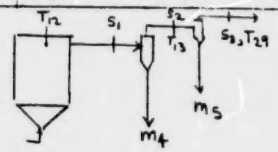
5.3 Performance of Standard Cyclones as Primary Collectors

Table 9 summarizes data for the series of runs with standard cyclone No. 1 (Figure 3) as the primary collector, and with either standard cyclone No. 2 or the vortex pocket cyclone as the secondary collector. For the primary cyclone, the inlet velocity varied from 14.6 to 19 m/s at temperatures in the range 740 to 910°C. Measured inlet loadings ranged up to 0.2 kg solids/kg gas (200,000ppmw) as recycle of solids from the primary cyclone was generally used. Mass balance calculations (making assumptions that all H and O in the coal was gasified) showed inlet values reaching a maximum of 0.35 kg solids/kg gas. These values are in the range of the 0.28 kg solids/kg gas quoted from literature references in section 1.0. Measured outlet concentrations from the primary cyclone averaged 0.0025 kg solids/kg gas (or 2500ppmw). This value is close to that suggested by Oakley and Reed[7] for a fluid bed gasifier after cyclone (2000ppmw), but lower than the 3000ppmw reported by Hozumi et al.[8] for two stages of cyclone separation. The measured values suggest average primary cyclone efficiencies of about 98.7%. These values have been compared with the efficiencies predicted by the equations of Leith which are listed below in Table 11. The agreement is quite satisfactory giving an average for the larger particle sizes reported of 0.98 over runs 143 to 147.

The composition of the primary cyclone char is given below in Tables 10 and 11.

TABLE 9: Solids Loading Results - Standard Cyclone No. 1 as Primary Separator

RUN NO.	DRY GAS FLOW (m^3/h) 15°C	WET GAS FLOW (kg/h)	VOL. LOADING (g/m^3 dry gas at 15°C)			MASS LOADING (g/kg wet gas)			PRIMARY[1] CYCLONE		MASS FLOWS (kg/h)		SECONDARY[2] CYCLONE		SAMPLE POINT S_2		SAMPLE POINT S_3		CYCLONE[3] EFFICIENCY (%)
			S_1	S_2	S_3	W_1	W_2	W_3	T_{in} (°C)	V_{in} (m/s)	m_4	m_5	T_{in} (°C)	V_{in} m/s	V_{pipe} (m/s)	V_s	V_{pipe} (m/s)	V_s	
134	39	98	450	30.6	—	179	12.2		800	17.2	16.3		525	20.2	22.	10.4	—	—	93.2
135	50	91	400	2.1	—	220	1.1		841	16.6	50.0		516	19.1	22.	7.6	—	—	99.8
136	64	89	192	1.7	—	139	1.3		748	14.6	43.1		477	20.6	22.6	13	—	—	99.8
137	76	95	57	4.0	—	46	3.2		737	17.3	17.6		458	19.8	20.9	—	—	—	98.3
138	64	91	65	3.7	—	46	2.6		907	19.2	19.9		490	21.0	24.2	—	—	—	98.8
141	41	91	216	—	0.65	98	—	0.30	820	17.0	23.0	3.86	404	17.4	22.3	—	20.5	52	—
142	36	88	241	—	1.05	99	—	0.43	814	17.7	24.4	0.65	440	18.3	22.9	—	20.7	50	—
143	30	93	192	16.5	0.4	63	5.4	0.13	910	18.7	21.7	0.50	450	55.5	25.7	103.	21.9	80.6	97.7
144	23	90	312	7.7	1.05	82	2.0	0.39	842	18.3	7.2	0.16	426	51.8	23.6	59.5	21.7	59.5	97.6
145	33/42	86/93	216	2.6	0.74	89	1.1	0.30	866	18.8	24.4	0.54	450	55.6	24.5	28.6	22.9	28.6	99.7
146	30	85	172	5.6	0.46	60	2.0	0.16	898	17.7	21.3	0.48	462	50.1	22.8	105.	20.6	78.7	99.2
147	29	88	—	5.5	—		1.8	—	788	16.6	26.0	0.76	495	55.4	25.0	55.8	20.5	—	99.4
149	24	66	—	10.9	—		4.0	—	833	14.4	21.0	0.63	418	41.6	14.4	81.3	12.5	—	98.8



(1) BASED ON T_{12} (2) BASED ON $(T_{13} + T_{29})/2$ (3) RUNS 134-138 % FOR BOTH CYCLONES 1 + 2. OTHERS, FOR PRIM. CYCLONE

TABLE 10: Average Composition of Primary Cyclone Char - Std. Cyclone No. 1

Feed	Primary Cyclone % Carbon	Product Composition % Ash	No of Runs
Suncrude Coke	62.7	37.3	3
Suncore Coke	90.7	9.3	2
Anthracite	61.0	39.0	8

TABLE 11: Composition and Size for Conventional Primary Cyclone

No	Feedstock	Recycle from Primary	Cyclone Char %C	Cyclone Char % Ash	Diameter of Particles (μm) Collected Sieve	Diameter of Particles (μm) Collected Coulter Counter	Passing	$\bar{\eta}$ Calc. Coulter
	Syncrude	Yes	63.8	36.2	89/91*	68/69	47/24	.999
	Syncrude	Yes	62.3	37.7	70/73	70/85		.999**
	Suncor	Yes	85.2	14.8				
	Suncor	Yes	96.2	3.8	91/59	53/50		.999
	Mt. Klappan	Yes	62.2	37.8	87	47		.997
	"	Yes	67.7	32.3		22		.963
	"	No	63.2	36.8		46		.997
	"	Yes	63.9	36.1		17		.945
	"	Yes	53.8	46.2		115		.999
	"	No	62.2	37.8		21		.969
	"	Yes	61.0	39.0		14/18	11	.953
	"	Yes	59.1	40.9		38/31	13	.995
	"	Yes	60.6	39.4		27/26	12	.983
	Mt. Klappan	Yes	57.3	42.7				

rst value refers to cyclone product, second value refers to dust drum overflow (see g. 2). ** Based on larger of two diameters, and calculated from eq'n of Leith.

The char analyses are in slightly higher in ash than the 30% figure quoted by Oakley and Reed[7]. No effect of recycle on char composition was evident.

Particle size results for the primary char are summarized in Table 11. For the relatively coarse material collected in the primary cyclone, sieving and Coulter counter methods were used. A reasonable agreement is seen between the two methods, however at mean diameters of 50μm and less, dry screening is to be avoided. The magnitude of the mean particle size collected gives an indication as to why the cyclone efficiencies are high. The particles collected from the oil sands runs averaged about 65μm in diameter, whereas those from the anthracite were smaller at a diameter of about 35μm.

Photomicrographs of these particles collected by the conventional primary cyclone are shown in Figure 8. Figure 8a for Run 145 shows particles to about 50μm, and very little below 10μm. Figure 8b shows Suncor coke particles from the primary cyclone of Run 138.

Photomicrographs of particles which have passed through the standard primary cyclone are shown in Figure 9. Figure 9a shows a typical example from Run 147. While there are a lot of small pieces in the range of a few microns, large particles of the order of 12μm are also evident.

5.4 Standard Cyclone Versus Vortex Pocket Cyclone As A Secondary Collector

A comparison of standard cyclone No. 2 and the vortex pocket cyclone as secondary devices following a standard primary cyclone was not too revealing. From the data of Table 9, it must be noted that the velocity entering the secondary is much higher in the case of the vortex pocket cyclone, and the barrel diameter is also considerably smaller at 10.2 cm than

the 19.4 cm of the conventional cyclone with which it is being compared. Mass flows collected as secondary cyclones cannot be directly compared since the configuration used in Runs 134-140 (Figure 2) resulted in primary and secondary collected particles being mixed. Thus, reliance on the gas sampling measurements alone was necessary, but again the data was insufficient to detect any differences in a statistical sense. Measured values of the exit gas loading from the two cyclones in series gave an average of 370ppmw for two runs with the standard secondary versus 250ppmw over four runs for the vortex pocket collector as secondary.

Secondary cyclone products analyze about 65.7% ash and 34.3% carbon (Table 12), as compared to 40% ash and 60% carbon for the primary products. This is a significant difference, and no doubt represents collection of particles of increased specific gravity in the secondary cyclone. Examination of the scattered particle size data suggests that the material passing the secondary cyclones is of the same mean size as that being collected i.e. ≈ 5-15μm. The material passing may be of slightly different composition than that collected however. For the vortex pocket cyclone, there is a suggestion that the particles collected in the pocket are slightly larger than those collected in the main leg. Efficiencies of the secondary cyclones were calculated using the collected mass rate and the solids passing the cyclone as determined from sample point 3 downstream of the secondary cyclone. For Standard Cyclone 2 as the secondary, calculations from the equation of Leith showed expected efficiency of 91.7% for d_p = 10μm and 98.4% for d_p = 20μm. Thus the measured values for Runs 141 and 142 given in Table 12 appear to be within the expected range.

TABLE 12: Collection Rate, Composition and Particle Size for Secondary Cyclones

Run	Secondary Cyclone	Collection Rate (kg/h)	Char Composition (% Wt.) C	Ash	Mean particle Diam. (μm) Collected	Passing	Cyclone Efficiency (%)
141	STD - 2	1.16	35.2	64.8	5.5	4.1/20.6	97.7
142	STD - 2	0.65	37.5	62.5	4.3	6.0/10.0	94.5
143	Vortex Pocket	0.50	35.4	64.6	4.6/3.9*		97.6
144		0.16	18.3	81.7	14.5	12(?)	81.7
145		0.54	37.4	62.6	16.9/10	15	95.3
146		0.48	32.8	67.2	9.0/12	12	97.2
147		0.76	33.3	66.7	0.4/13		97.2
149	Vortex Pocket	0.63	34.9	65.1			

Anthracite Feedstock. * First value from vortex pocket, second from main leg.

TABLE 13: Vortex Pocket Cyclone As Secondary Collector

Run	Vortex Pocket Collection Mass Rate (kg/h)	%C	%Ash	Main Leg Collection Mass Rate (kg/h)	%C	%Ash
143	0.11	43.0	57.0	0.39	33.3	66.7
144	0.014			0.142	18.4	81.6
145	0.098	46.8	53.2	0.442	35.3	64.7
146	0.085	37.9	62.1	0.390	30.8	69.2
147	0.141	34.1	65.9	0.618	33.0	67.0
149	0.132	40.3	59.7	0.302	34.9	65.1

$$\eta = \frac{1}{1 + \frac{GS_3}{M_{collected}}} \tag{7}$$

The average for six runs is 96.6%. The average particle size collected is 9.4μm. Where the vortex pocket cyclone was used as the secondary cyclone, the mass flows in the main leg exceeded those in the pocket by a factor of about five (Table 13). The main leg particles are slightly enriched in ash compared to the pocket-collected particles.

Samples of materials collected in the secondary cyclone are shown in Figures 10. Figure 10a shows a sample at high magnification from Run 143 where the vortex cyclone was used as the secondary collector. Particle sizes are generally below 10μm with many particles visible in the 4-5μm size range. Some larger particles are seen to have escaped the primary cyclone in Run 145, where in Figure 10b, which was taken at lower magnification, a few particles of about 50μm are evident.

5.5 Vortex Pocket Cyclone as Primary Collector

The vortex pocket cyclone was used as a primary cyclone for eleven runs. Results are summarized in Tables 14 and 15. The inlet loading was generally well below the value tested for the conventional cyclone in most runs, being in the range 0.04 to 0.141 kg solids/kg gas, as no recycle was used. The inlet velocity was extremely high because of the small entrance area. Efficiencies averaged over 99.3% for 9 of the 11 runs. Outlet loadings for eight of eleven runs averaged 390ppmw. This value is about a factor six below the mean value for the standard cyclone No. 1. Pressure drops are undoubtedly lower with the standard unit.

Some interesting features in terms of collection rates and collected solids composition were noted. (Table 15). When used as a primary cyclone the ratio of mass flow collected in the pocket to that of the main cyclone leg averaged 7.3. Furthermore, the ash content of the vortex pocket solids was 34% compared to 67% ash in the main leg of the cyclone.

It was noted that the presence of a second conventional cyclone downstream of the vortex pocket cyclone (Runs 150 to 152) collected virtually no further material i.e. <0.5% of that collected in the primary was found in standard cyclone No. 2.

Photomicrographs of particles collected with the vortex cyclone as the primary collector are shown in Figure 11. The solids are roughly comparable to those collected by the conventional primary cyclone, and contain a large number of fine particles of the order of 5μm. Because of the high superficial velocities in the gasifier, much coarser material is carried over. Figure 12 shows particles escaping the vortex cyclone. A few very large (300μm) particles are evident.

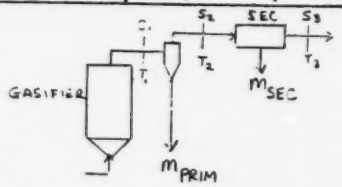
5.6 Granular Bed Filter

The granular bed filter was tested only as a secondary device, with the vortex pocket collector cyclone upstream as the primary separator. The configuration of the test equipment is shown in Figure 12. Because the reactor off-gases contain so much moisture it was necessary to build a

TABLE 14

Solids Loading Results for Vortex Pocket Cyclone as Primary Collector

RUN NO	DRY GAS FLOW (m^3/h) 15°C	WET GAS FLOW (kg/h)	VOL. LOADING (g/n-m^3 dry gas)			MASS LOADING (g/kg wet gas)			CYCLONE		MASS FLOWS (kg/h)		SAMPLE PT. S_2		SAMPLE PT. S_3		W_1 BY MASS BALANCE (g/kg)	EFFICIENCY OF PRIMARY CYCLONE (%)
			S_1	S_2	S_3	W_1	W_2	W_3	T_{in} (°C)	V_{in} (m/s)	PRIM.	SEC.	V (m/s)	V_s	V (m/s)	V_s		
150	65.6	91.6		0.91	0.42		0.65	0.30	730	62	14.39	0.066	61.3	56.5	37.4	67.2	87	99.6
151	162	198	42.6	0.53	0.46	34.8	0.43	0.32	720	132	11.35	0.076	55	84	37	90	57	99.3
152	121	142	140	0.27	3.2	88	0.17	2.0	730	125	13.67	—	50	84	36	58	71	99.7
153	68	103	47	0.12	0.67	37	0.10	0.53	775	84	4.58	0.0081	28	85	22	72	84	99.8
154	118	169	151	0.72	0.50	103	0.49	0.34	770	126	10.04	0.0488	41	79	35	72	59	99.2
155	121	198		0.74	0.43/0.29		0.45	0.26	863	156	11.95	0.0122	51	104	42	82	60	99.2
156	92	120	78	0.59	0.65	60	0.41	0.65	676	74	7.34	0.0093	23	61	21	49	73	99.4
157	92	117	82	16.5	0.80	64	4.1	0.62	680	88	7.73	0.1556	27	22	23	47	76	83.7
158	55	97	5.17	28.2	5.48	2.7(?)	14.9	2.89	723	79	5.34	0.0416	24	60	23	62		77.6
159	113	142		0.39	0.65		0.34	0.56	663	82	7.05	0.0813	26	45	24	30	109	99.4
160	94	143		0.24	0.22		0.16	1.05	743	98	6.26	0.043	30	54	25	36	51	99.6



RUNS 150-2 STD CYC. #2 AS SECONDARY
153-60 GRAN. BED " "

TABLE 15

Collection Rate, Composition and particle Size for Vortex Pocket Cyclone as Primary Collector

RUN NO.	FEED STOCK*	VORTEX POCKET LEG				MAIN CYCLONE LEG			CYCLONE EFFICIENCY (%)
		MASS RATE (kg/h)	% C	% ASH	d_p ** (μm)	MASS RATE (kg/h)	% C	% ASH	
150	M-K	12.95	65.6	34.4	42.3	1.48	33.5	66.5	99.6
151	M-K	9.85	56.1	43.9	53.8	1.52	32.5	67.5	99.3
152	M-K	12.33	67.9	32.1	51.9	1.32	30.1	69.9	99.7
153	HV	4.06	55.5	44.5	63.1	0.53	41.3	58.7	99.8
154	HV	8.72	59.0	41.0	65.8	1.35	32.1	67.9	99.2
155	HV	10.48	66.7	33.3		1.49	40.1	59.9	99.2
156	HV	5.83	62.3	37.7		1.21	38.3	61.7	99.4
157	HV	6.50	70.8	29.2	75.7	1.23	34.8	65.2	83.7
158	M-K	4.70	71.9	28.1	70.4	0.64	30.7	69.3	77.6
159	M-K	6.27	75.0	25.0	70.6	0.79	30.2	69.8	99.4
160	M-K	5.67	74.0	26.0	66.9	0.59	24.5	75.5	99.6

* M-K MT. KLAPPAN ANTHRACITE, HV HIGHVALE COAL

** BY SIEVING PLUS COULTER COUNTER; SAUTER MEAN VALUE

pre-heater for the granules such that they would not become coated with condensed water during start up. In the configuration shown the gas flows upwards through the bed of particles. Because of the efficiency of the vortex cyclone the inlet solids loadings to the granular bed were generally low. Table 16 outlines conditions of the runs and results. Eight runs were carried out using two feedstocks, three levels of velocity, two active bed depths and three different collector materials.

The initial experiments utilized a gravel of size distribution shown in Table 17, and mean d_p of 2.2 mm. A single test was undertaken with silica particles of a smaller particle size of 1.2 mm, however this material was partially blown out of the bed, and subsequent tests were done with a coarser material. The last four trials used a dolomite of size distribution given in Table 17 and a mean diameter of 2.2 mm.

Because of the conical geometry of the lower section of the bed (which was constructed in this manner to permit future continuous discharge of solids), a portion of the bed will be inactive. From the geometry shown in Figure 5, the inactive bed depth is assumed to be that below the gas outlet, although some of the bed volume no doubt will collect particles. The conical section will also result in a change in velocity of the gas with bed depth. For calculation purposes the cross-sectional area available for flow was taken to be the average of that at the inlet of the active region (0.0893 m^2), and that at the top of the bed. The bed top fell within the cylindrical section of the granular bed for all but the initial run.

Efficiencies of the granular bed filter were calculated in three different ways, and the results averaged where appropriate. This was done to attempt to average out errors in sampling. The granular bed was operated in the semi batch mode in which solids built up over time on the granules.

TABLE 16 Summary of Granular Bed Tests

N	FEED-STOCK	GRANULAR BED MATERIAL	BED WT. (kg)	BED VOL. (m^3)	ACTIVE BED VOL m^3	ACTIVE BED DEPTH m	AVG. CROSS-SECTN AREA m^2	WET GAS FLOW at 500°C (m^3/h)	U_0 GRANULAR BED at 500°C (m/s)	MASS LOADING (ppmw) IN	MASS LOADING (ppmw) OUT	MASS FLOWS OF SOLIDS (kg/h) IN	MASS FLOWS OF SOLIDS (kg/h) OUT	MASS FLOWS OF SOLIDS (kg/h) COLL	EFFICIENCY η_1	EFFICIENCY η_2	EFFICIENCY η_3	EFFICIENCY $\bar{\eta}$	TEMP** (°C)
3	H-V	GRAVEL	14.61	.0297	0.021	0.148	0.140	288	0.57	550(est)	530	.0083	.0464	.0081		.15		.15	447
4	H-V	"	83.85	.0559	0.0472	0.268	0.185	418	0.63	494	340	.085	.0585	.0488	.574	.455	.68	.57	453
5	H-V	"	65.60	.0437	0.0350	0.247	0.170	495	0.81	450	260	.089	.0515	.0122	.137	.192	.42	.25	522
6	H-V	SILICA 10	80.81*	.0608	0.0521	0.285	0.185	281	0.42	410	650	.049	.078	.0093	.189	.107		.15	392
7	H-V	DOLOMITE	84.9	.0578	0.0491	0.274	0.185	281	0.42	4100(?)	615	.48	.0725	.1556	.324	.682	.15	.39	364
8	M1-K	"	81.3	.0553	0.0465	0.266	0.185	295	0.44	14,900	2890/5580	155	.300	.0416	.027	.122	.81	.32	396
9	M1-K	"	76.6	.0521	0.0434	0.254	0.185	295	0.44	340	560	0.044	.0717	.0813	>1	.531	÷	.53	411
0	M1-K	"	72.4	.0493	0.0406	0.229	0.181	352	0.54	160	140	0.023	.020	.043	>1	.683	.13	.41	404

* BED PARTIALLY BLOWN OUT DURING EXPT. ** AVG INLET & OUTLET TO G.B.F.

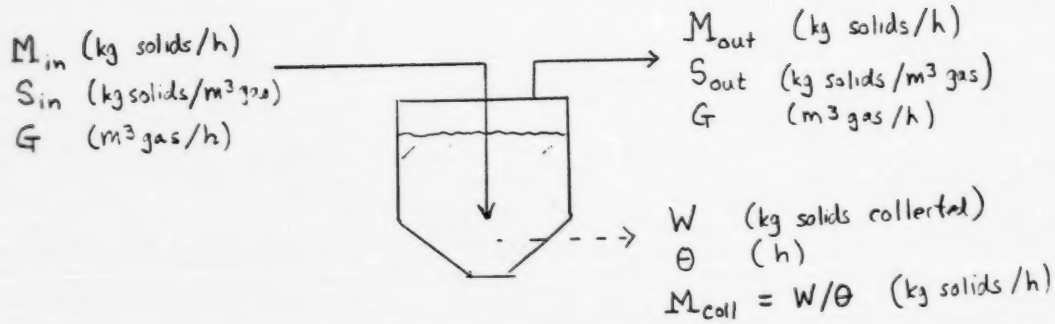
TABLE 17: Size Distribution of Granular Bed Collector Materials

Avg. Screen Aperture (mm)	Gravel Wt. Fraction (x_i)	Avg. Screen Aperture (mm)	Silica 10 Wt. Fraction (x_i)	Dolomite Wt. Fraction (x_i)
4.06	0.0613	3.57	0.0127	0.2681
2.87	0.5353	2.03	0.2627	0.6618
2.18	0.2063	1.435	0.4169	0.0691
1.85	0.0887	1.02	0.1535	0.0010
1.55	0.0353	0.718	0.0925	
1.29	0.0247	0.475	0.0499	
1.02	0.0305	0.283	0.0086	
0.718	0.0101	0.18	0.0003	
0.475	0.0042			
0.302	0.0036			
0.100	0.0000			
$d_p = \frac{1}{\sum_i \frac{x_i}{d_i}}$ (mm)	2.22		1.19	2.22

TABLE 18: Composition of Particles Collected in Granular Bed

Run	Mt. Klappan	Collector Material	Analysis of Collected Material (Wt.%) C	H	Ash*
155	Highvale	Gravel	32.1	0.94	67.0
156	Highvale	Silica Sand	14.1	0.87	85
157	Highvale	Dolomite	19.0	<0.5	81
158	Mt. Klappan	Dolomite	16.0	<0.5	83.7
159	Mt. Klappan	Dolomite	13.2	<0.5	86.5
160	Mt. Klappan	Dolomite	14.0	0.5	85.6

* by difference



The mass of collected solids was determined at the end of each run by floating the collected solids from the bed material (Appendix D). Thus, the hourly mass collected, M_{coll}, was found as mass collected/hours of operation. Efficiencies were then evaluated as

$$\eta_1 = M_{coll}/M_{in} = M_{coll}/S_{in}G \tag{8}$$

$$\eta_2 = M_{coll}/(M_{coll} + M_{out}) = \frac{1}{1 + S_{out}\ G/M_{coll}} \tag{9}$$

and $$\eta_3 = (M_{in} - M_{out})/M_{in} = 1 - \frac{S_{out}}{S_{in}} \tag{10}$$

the values of S_{in} and S_{out} were determined by sampling the gas stream as before.

Some difficulties arose with respect to the gas loading results. As can be seen in Table 16, outlet loadings were in three cases higher than inlet loadings, and for some runs the results were inexplicably high. The increase in loadings may have been due to dolomite dust on the surface of the dolomite, or just the result of errors. The calculated exit loadings for this configuration averaged 300ppmw for two runs with gravel at the deepest

active bed depth of 0.26 m. Examination of the data of Table 16 suggests bed collection efficiency increases with bed depth (Run 153 vs 154), and increasing velocity has no clear effect. Table 18 gives the analysis of collected particles which were subsequently floated off the surface of the bed as described in Appendix D. The collected material from Run 155 with clean gravel as the bed material was about 67% ash, whereas for the remainder of the five runs the collected material averaged about 84.5% ash. This result, combined with previous data clearly shows the increasing ash content in going from primary collection to secondary. Thus Tables 11 and 15 show primary collection gives material of about 65% carbon, Table 12 shows secondary cyclones yielding material of about 35% carbon and Table 18 shows the granular bed to be collecting material of about 15% carbon.

The average collection efficiency over the six runs at the deeper bed depth is about 37% (with a considerable scatter). If we take the average exit loading for the vortex pocket collector as 390ppmw (Section 5.5), then the outlet loading is expected to be about 250ppmw. This overall average value is slightly high, but still in the region of the 200ppmw listed in Section 1.0 as being the turbine tolerance limit as far as projected total loadings are concerned. However, this value was not reached in the present experiments.

The scattered and relatively low collection efficiencies of the granular bed was a cause for concern, and it was considered desireable to check the measured range of values against predictions of established mathematical models for granular beds applied to the conditions of the tests. The following comparison was made:

The predicted granular bed collection efficiencies were computed for conditions of several runs using equations of Thambimuthu et al.[20], Schmidt et al.[21], and Paretsky et al.[22]. These models give different expressions for the single particle collection efficiency, and then combine them into a total bed collection efficiency. For the particles of the size range of interest here i.e. $0.5\mu m < d_p < 5\mu m$, the major mechanism of importance particularly for the larger particles is collection by inertial deposition. Each of these models uses a different term for the inertial term. Following Thambimuthu et al.,

$$E_{particle} = \underbrace{E_D}_{\text{diffusion term}} + \underbrace{E_I}_{\text{inertial impaction term}} + \underbrace{E_G}_{\text{gravity term}} + \underbrace{E_{DI}}_{\text{direct interception term}} \tag{11}$$

and the overall bed collection efficiency is given by

$$E_{bed} = 1 - \exp\left[-1.5\,\frac{(1-\varepsilon)}{\varepsilon}\cdot\frac{L_b}{d_c}\cdot E_{particle}\right] \tag{12}$$

where ε = bed voidage

L_b = active bed depth

d_c = diameter of collector particle

In subsequent calculations the following equations have been used:

$$E_D = \frac{4.36}{\varepsilon}\left(\frac{D}{U_o d_c}\right)^{2/3} \tag{13}$$

where D is the particle diffusivity, given by the Stokes-Einstein equation

$(k_B T/3\pi\mu d_p)$

U_o = superficial velocity through the packed bed

$$E_G = 0.0375\ (V_T/U_o)^{0.5} \tag{14}$$

$$E_{DI} = \frac{6.3}{\varepsilon^{2.4}} \cdot (d_p/d_c)^2 \tag{15}$$

Equations (13), (14) and (15) are due to Paretsky, and quoted by Thambimuthu and others. For the inertial impaction term three different equations were used:

$$E_I = \left(\frac{Stk}{Stk + .062\varepsilon}\right)^3 \qquad \text{Thambimuthu} \tag{16}$$

$$E_I = \frac{\pi}{0.4(1-\varepsilon)(3.75)}\ Stk \qquad \text{Schmidt} \tag{17}$$

$$E_I = 2(Stk)^{1.13} \qquad \text{Paretsky} \tag{18}$$

where Stk = Stokes No. = $\rho_p d_p^2 U_o/9\mu d_c$

and ρ_p = particle density and μ = gas viscosity

Figure 14a shows the range of prediction of the three equations, and suggests collection efficiencies below 50% for particles below 1 - 2 microns.

Figure 14 shows typical predictions for the gas composition and bed geometry of Runs 153, 154 and 155. Here $\rho_p = 1500\ kg/m^3$. It is clear that for the range of particle size of interest, the expected bed efficiency is a very strong function of particle diameter. To put the current or future work on granular beds on a more sound basis, effort should be given to efficiencies integrated over the particle size distribution, and measurements made of particle densities. A correct description of the velocity field in the non-cylindrical granular bed collector should be derived and measured.

Photomicrographs of the loaded granular bed particles are shown as Figure 15. Figure 15a shows three granules of gravel at low magnification from Run 155. The fine particles attached to the surface are clearly visible. Micrograph 15b shows a close-up of the deposited char from Runs 153. Most deposited particles appear to be in the sub-micron size range, although a few agglomerates and particles up to 5μm appear in view. Micrograph 15C shows the detritus present on the surface of dolomite collector particle from Run 160.

5.7 Photomicrographs of Particles Not Collected

Samples of materials which bypass the collection system are shown in Figure 16. Figure 16a shows the dust load from Run 144 which has passed through a conventional primary cyclone and a vortex pocket secondary cyclone. A few particles in excess of 5μm are visible, but the majority appear to be about 3μm. Figures 16b and c shows samples from Highvale gasification (Run 157 and 156) which have passed through the vortex cyclone and the granular bed. Note that individual particles from Run 157 are all smaller than 5μm although there appears to be an agglomerate of pieces about 2μm x 4μm for Run 156 two large particles of about 15μm are present along with a myriad of small particles.

REFERENCES

1. Dellefield, R.J. and R.C. Bedick, "Evaluation of Three High Temperature Particle Control Devices for Coal Gasification", I. Chem. E. symp. Ser. No. 99, 467-482 (1986).

2. Rubow, L.N., "Technical and Economic Evaluation of Ten High Temperature High Pressure Particulate Cleanup Systems for PFBC", Fluid. Comb'n - Is it Achieving its promise, I. Chem. E., London (1984) Disc 42/378-396.

3. Theodore, L. and Buonivore, A.J., "Industrial Air Pollution Control Equipment for Particulates", CRC Press, Cleveland, Ohio, USA (1976).

4. Alexander, R. McK. Proc. Australas. Inst. Mining and Met. N.S. 152-3: 203 (1949) cited in Ref. 11.

5. Leith, D. see Ref. 11.

6. Wheeldon, J.M., G.K. Burnard, G.C. Snow and L. Svarovsky, "The Performance of Cyclones in the Off-Gas Path of a Pressurized Fluidised Bed Combustor", I. Chem. Symp. Ser. No. 99, p. 45-65 (1986).

7. Oakley, J.E. and G.P. Reed, "Behaviour of High Temp. Filter Materials in Hot Gasifier and Combustor Gas Atmospheres", ibid. p. 421-444.

8. Hozumi, S., J. Fukui, K. Kodama, K Kamei, T. Mori, N. Izumi and S. Kusada, "Gas Cleaning at High Temp: Present Stage of Granular Dust Removal Process Dev. for Gas Cleaning of Coal Gas", ibid. p. 385-399.

9. Makherjee, D.K., D. Mueller and O. Tassicker, "Flue Gas Cleanup Devices for PFBC Power Plants", Fluid Bed Comb'n Is it Achieving its Promise? I. Chem. E., London, Disc/44/408-419 (1984).

10. Watkinson, A.P., G. Cheng and R. Quan, "Gasification of Oil Sands Coke", DSS Contact Report OST 84-00055 (1986).

11. "Handbook of Powder Science and Technology", M.E. Fayed & L. Otten (EDS.) chapter 16 - Cyclones by D. Leith Van Nostrand, New York (1984).

12. Stairmand, C.J., "The Design and Performance of Cyclone Separators", Trans. I. Ch. E. 29, 356 (1951).

13. Swift, P., Steam and Heating Eng. 38, 453 (1969).

14. Billin, M., N. Syred and P. Sage, "Enhanced Collection Efficiency of New Concepts for Compact Cyclone Dust Separators for Hot Gas Cleanup", I. Chem. E., Symp. Ser. No. 99, p.17-30 (1986).

15. Sage, P.W., M.A. Wright and R.K. Clark, "Particulate Removal Developments for Industrial Coal-Fired Fluidised Bed Boilers", Fluid. Bed Comb'n - Is it Achieving its Promise?, I. Ch. E., London, Disc/43/397-407 (1984).

16. Syred, N., M. Biffin, S. Dolbeau, M. Wright and P. Sage, "Evolution of New Concepts for Compact Cyclone Dust Separators for Hot Gas Cleanup", I. Chem. E., Sym. Ser. No. 99, p. 17-30 (1986).

17. Shaw, G.D.H., J.G. Petrie and R.K. Dutkiewicz, "The Application of Moving Granular Panel Filter to Flue Gas Cleanup", ibid. 401-420 (1986).

18. Guillory, J.L., "Progress in High Temperature Moving Bed Granular Filter Development at Combustion Power Company", N.T.I.S. Ref. No. DE820/828/ (1982).

19. Badzioch, S., "Correction for Anisokinetic Sampling of Gas-borne Dust Particles", Jour. Inst. Fuel, p. 106-110, mar. (1960).

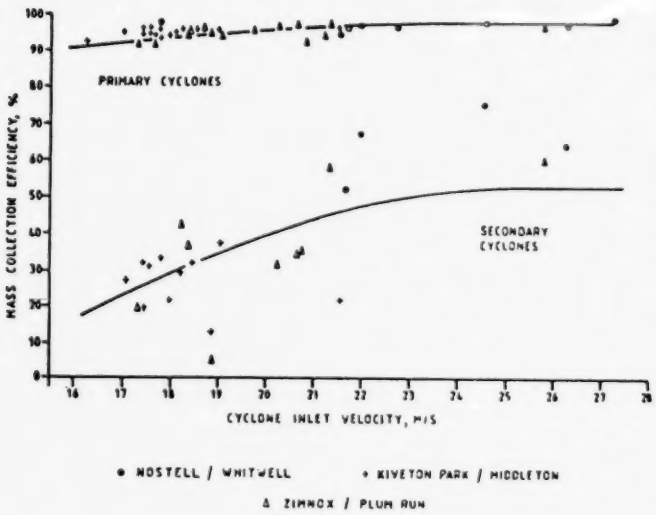
20. Thambimuthu, K.V., B.K. C. Tan and R. Clift, "Mechanism of Steady Filtration of Gases in Packed Beds", I. Chem. E. Symp. Ser., No. 59, 1:3/1 - 1:3/12 (1980).

21. Schmidt, E.W., J.A. gieseke, P. Gelfand, T.W. Lugar and D.A. Furlong, "Filtration Theory for Granular Beds", J. Air Poll. Control Assoc., 28, 143-146 (1978).

22. Paretsky, L., L. Theodore, R. Pfeffer and A.M. Squires, "Panel Bed Filters for Simultaneous Removal of Fly Ash and Sulphur Dioxide", J.Air Poll. Control Assoc., 21, 204 (1971).

23. Cadle, R.D., "The Measurement of Airborne Particles", Wiley, New York (1975).

FIGURE 1 - MASS COLLECTION EFFICIENCIES FOR PRIMARY AND SECONDARY CYCLONES VERSUS CYCLONE INLET VELOCITY



● NOSTELL / WHITWELL + KIVETON PARK / MIDDLETON

Δ ZIMMOX / PLUM RUN

Ref. 6

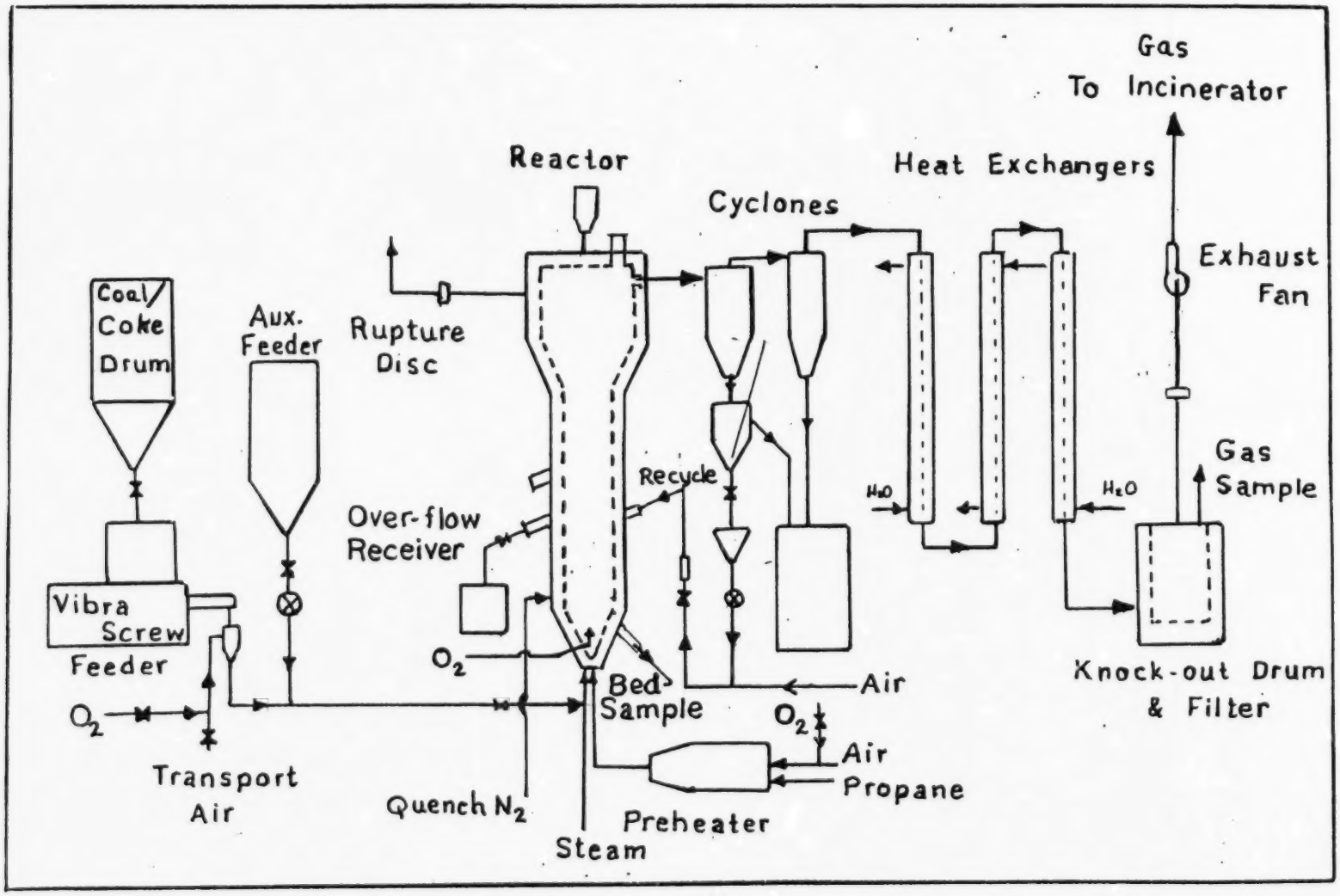
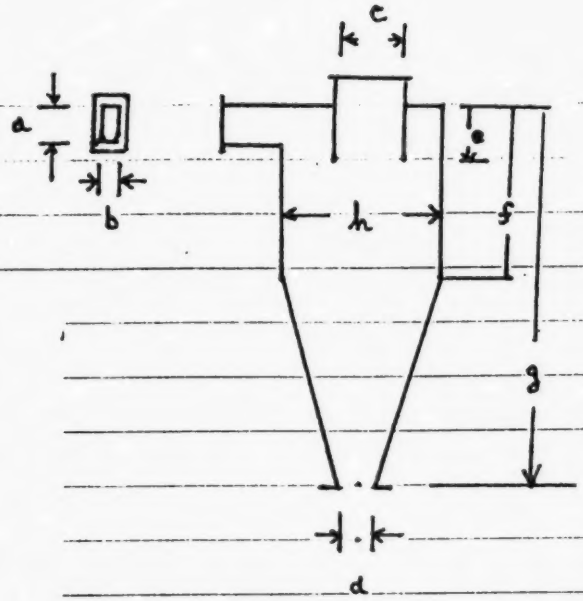


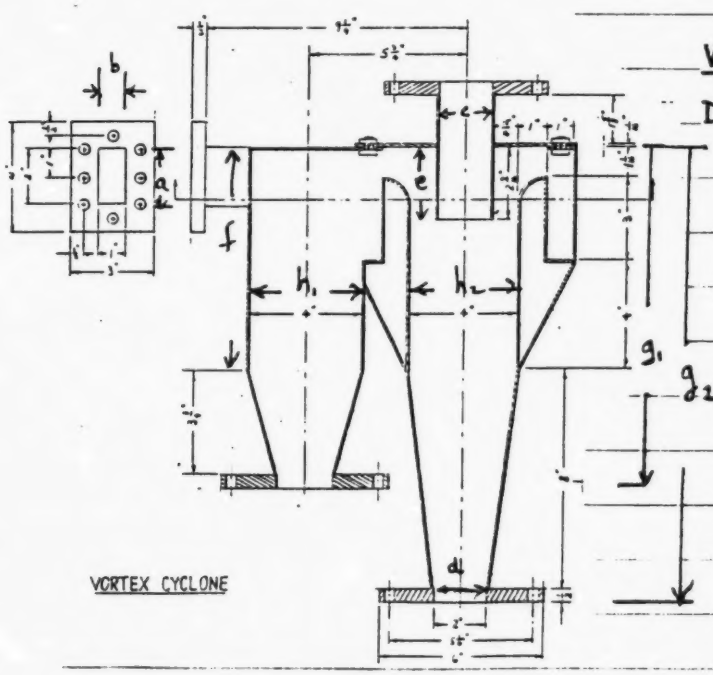
Figure 2 Schematic Diagram of Gasifier Pilot Plant

Figure 3 Standard Cyclones No. 1 and No. 2



STANDARD CYCLONE NO.	1	2.
DIMENSION a	12.07	8.38
b	4.45	4.06
c	7.62	7.62
d	7.62	7.62
e	12.07	9.5
f	27.3	26.7
g	61.0	74.9
h	20.6	19.4
Dimensions	(cm.)	

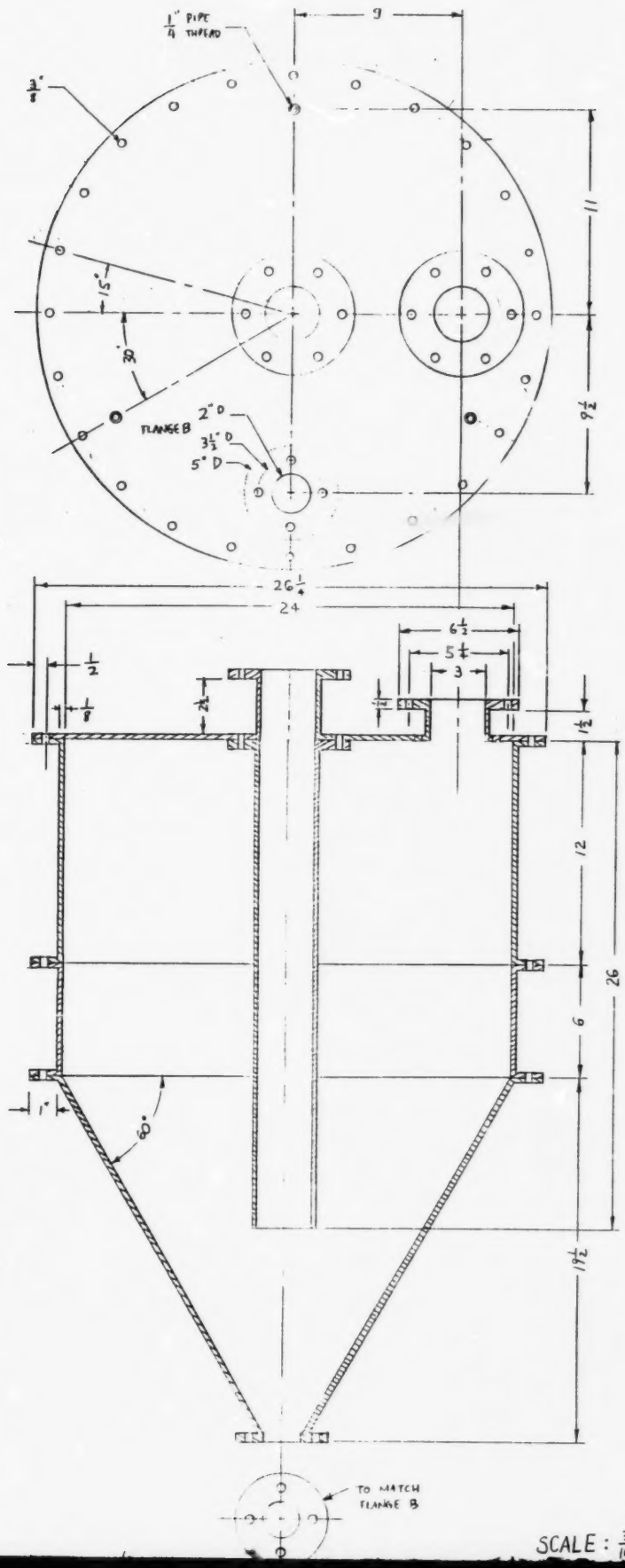
Figure 4 Vortex Pocket Cyclone



VORTEX	CYCLONE
DIMENSION a	5.1
b	2.54
c	5.1
d	5.1
e	6.51
f	20.5
g_1	30.0
g_2	40.8
h_1	10.2
h_2	10.2

* ALL DIMENSIONS IN CM.

Figure 5 Granular Bed Filter



GRANULAR BED
FILTRATION UNIT

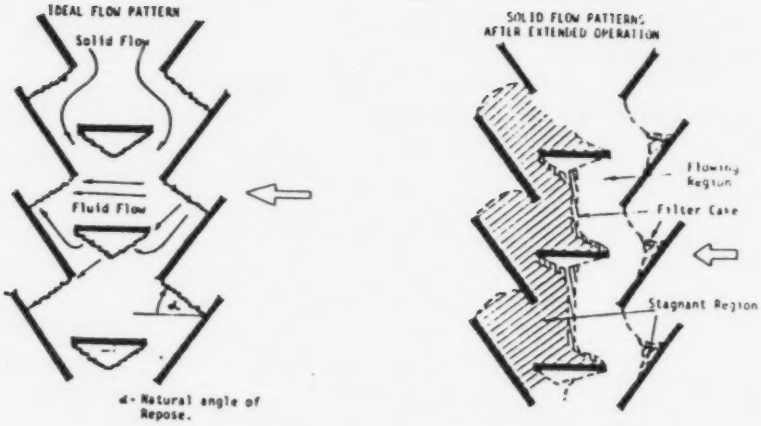
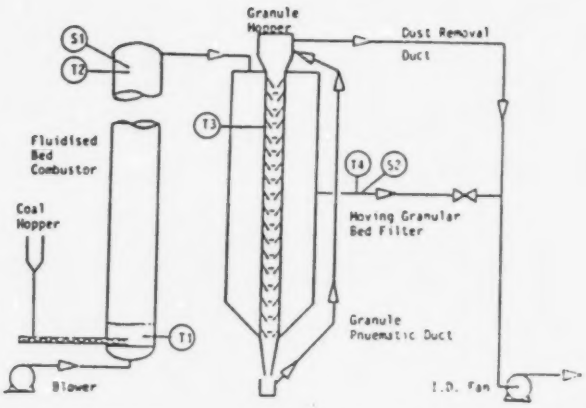


Figure 6a Cross Flow Granular Bed Filter of Shaw et al. (17)



FILTER GENERAL ARRANGEMENT

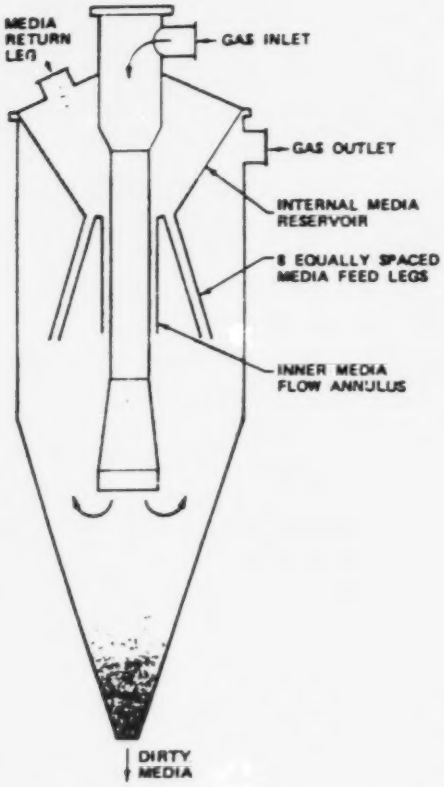


Figure 6b Countercurrent Granular Bed Filter of Combustion Power Co. (17)

Figure 1 Schematic of Moving Screenless Granular Bed Filter

Figure 7 Gas Sampling Train and Probe

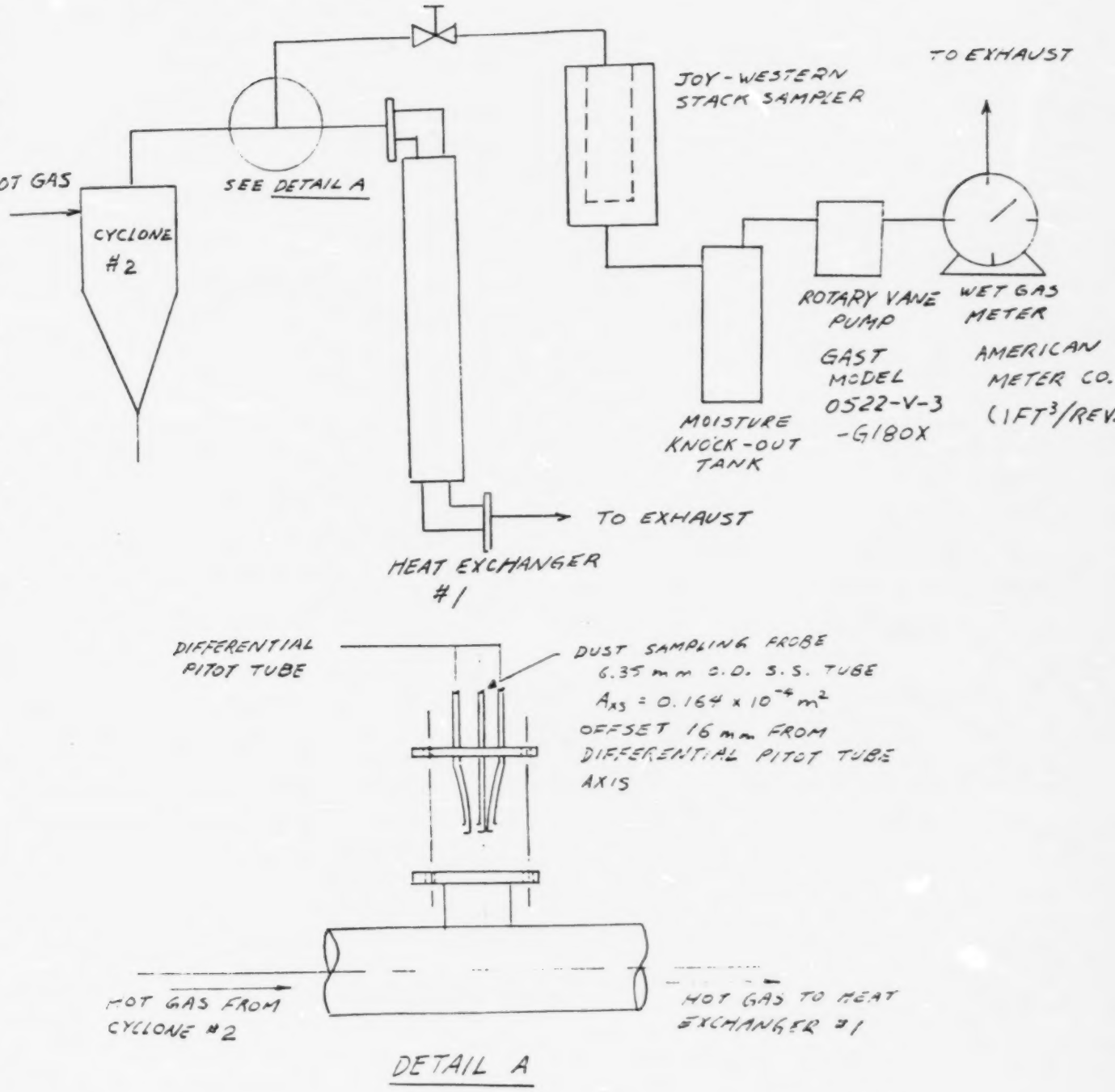
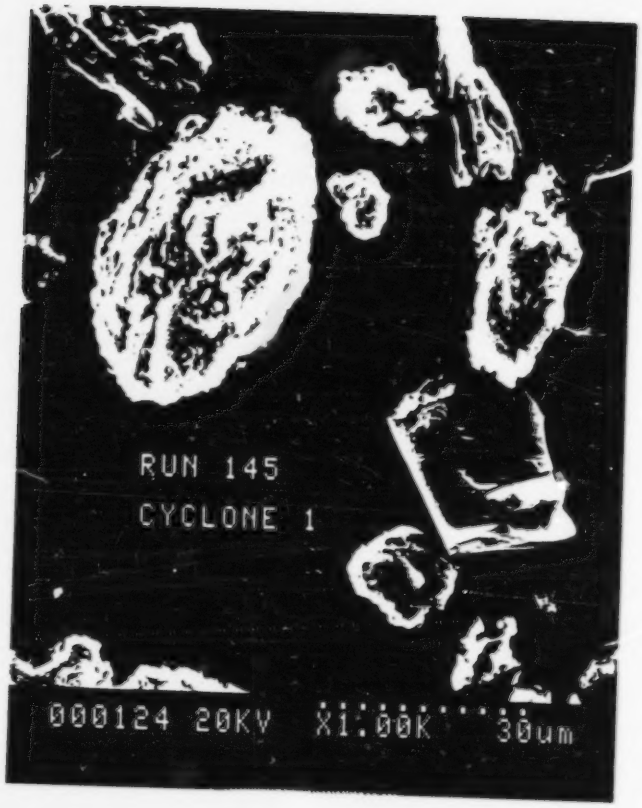


Figure 8
Photomicrographs of Particles Collected in Std. Cyclone No. 1 as Primary.

(a) Run 145 x 1000
Anthracite (U_o = 1.66 m/s)



(b) Run 138 x 100
Suncor (U_o = 1.6 m/s)

(52)

Figure 9 Particles Passing Through Standard No. 1 Primary Cyclone

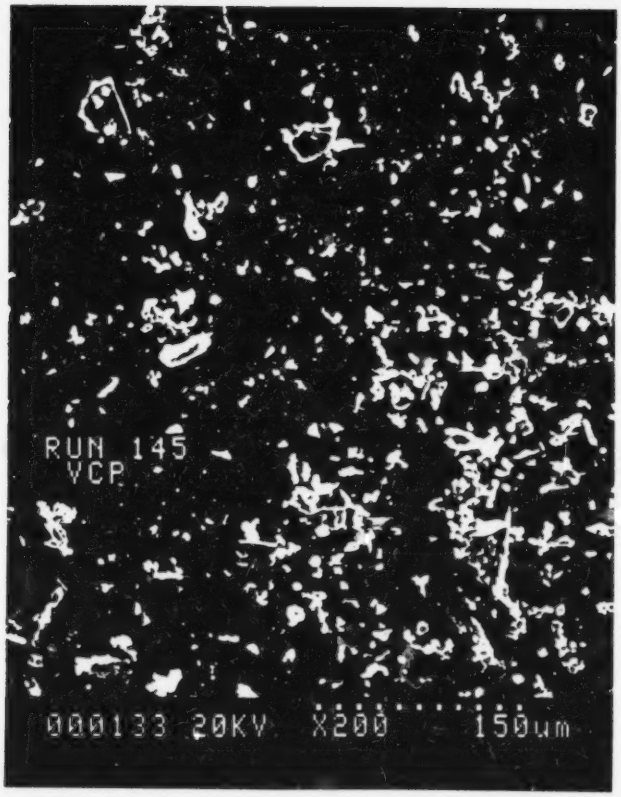
(a) Run 147 x 2500
Mt. Klappan Anthracite

(b) Run 146 x 1500
Mt. Klappan Anthracite

Figure 10 Particles Collected in Secondary Cyclone

(a) Run 143 x 2000
Mt. Klappan Anthracite
Vortex Pocket

(b) Run 145 x 200
Mt. Klappan Anthracite
Vortex Pocket



(a) Run 155 x 100
Highvale Coal
(U_o = 3.1 m/s)

(b) Run 157 x 100
Highvale Coal
(U_o = 1.7 m/s)

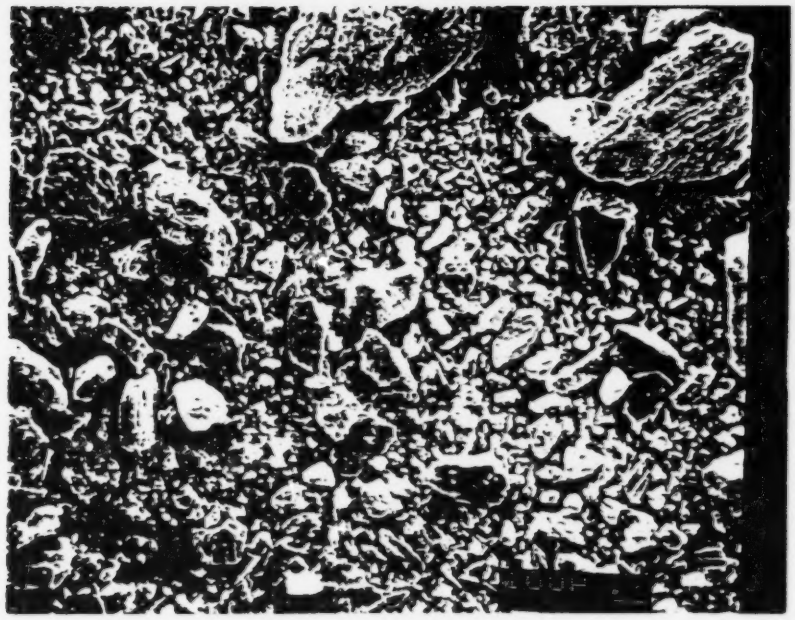
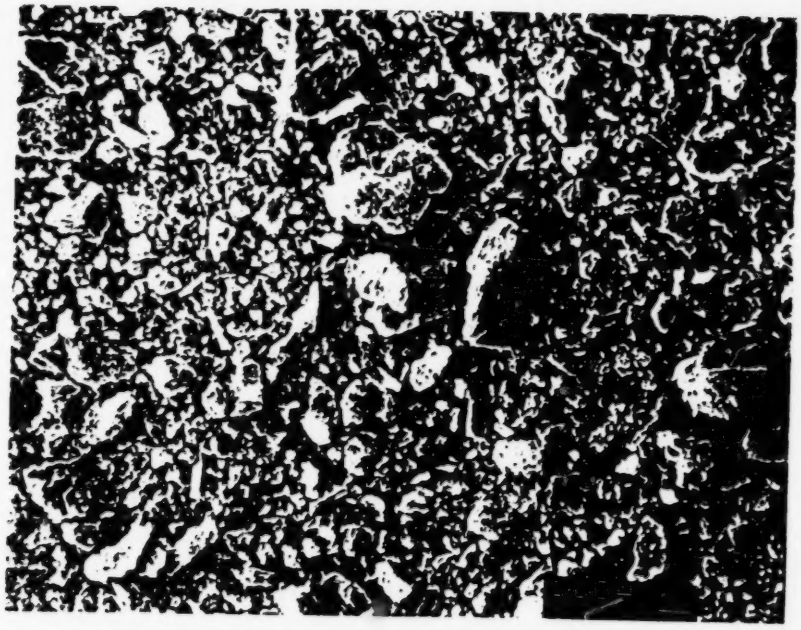
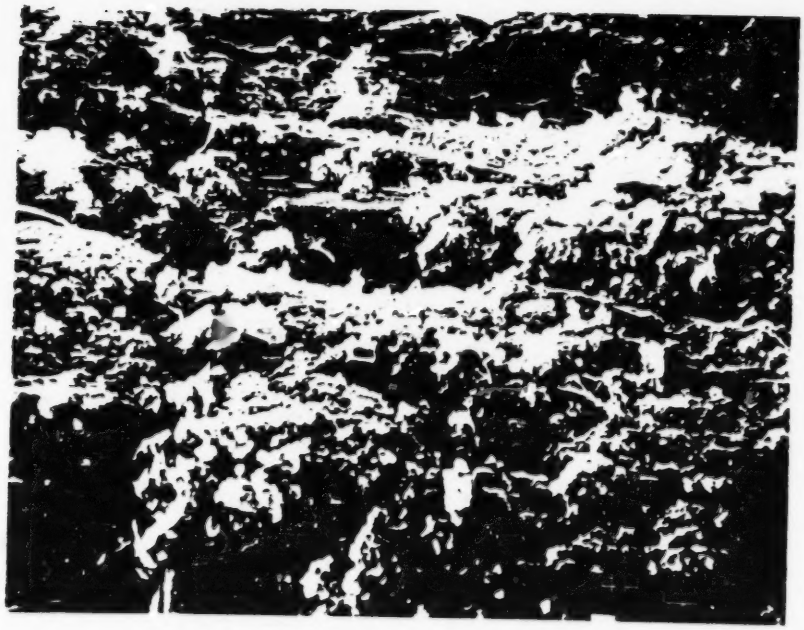


Figure 11 Particles Collected with Vortex Pocket cyclone as Primary

// (c) Run 160 x 100
Mt. Klappan
(U_o = 1.35 m/s)



(a) Run 157 x 100
Highvale



(b) Run 157 x 2500
Highvale

Figure 12 Particles Escaping Primary Vortex Pocket Cyclone

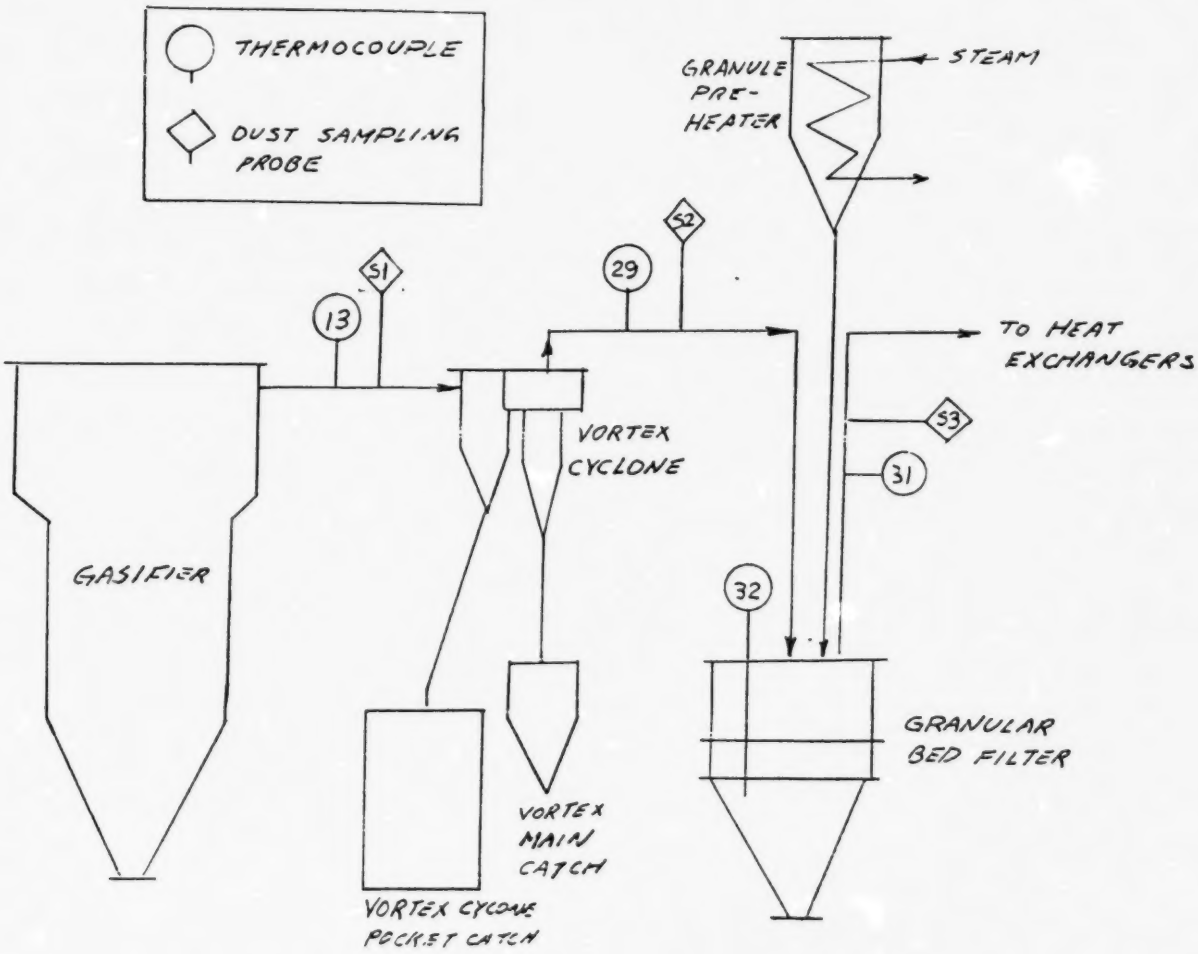


Figure 13 Layout of Apparatus for Granular Bed Experiments

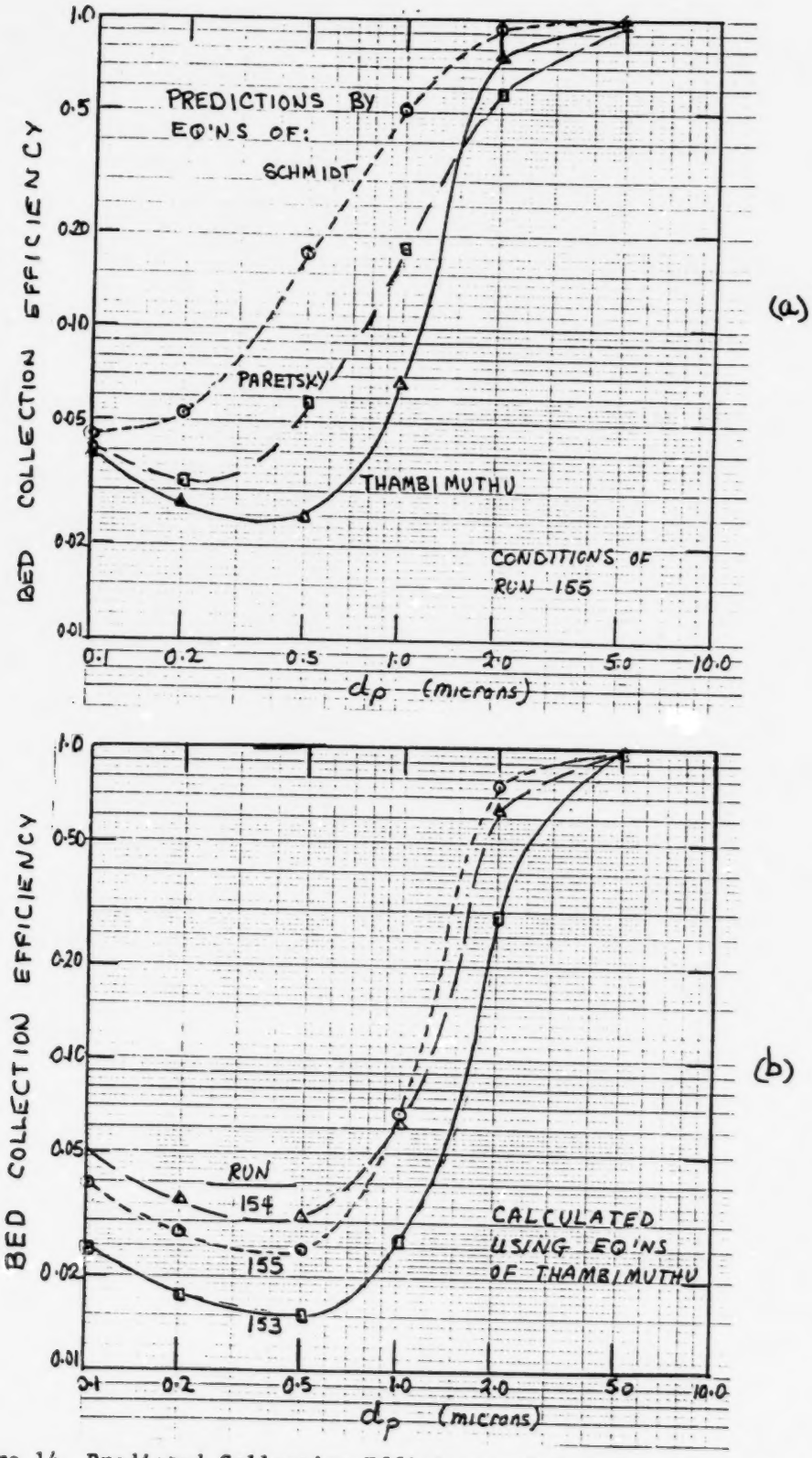


Figure 14 Predicted Collection Efficiency of Granular Bed

x 30

Figure 15 Granular Bed Collector Particles and Entrapped Dust

(a) Run 155 x 30
Gravel

(c) Run 153 x 2000
Highvale Dust on Gravel

(b) Run 160 x 1000
Anthracite Dust on Dolomite

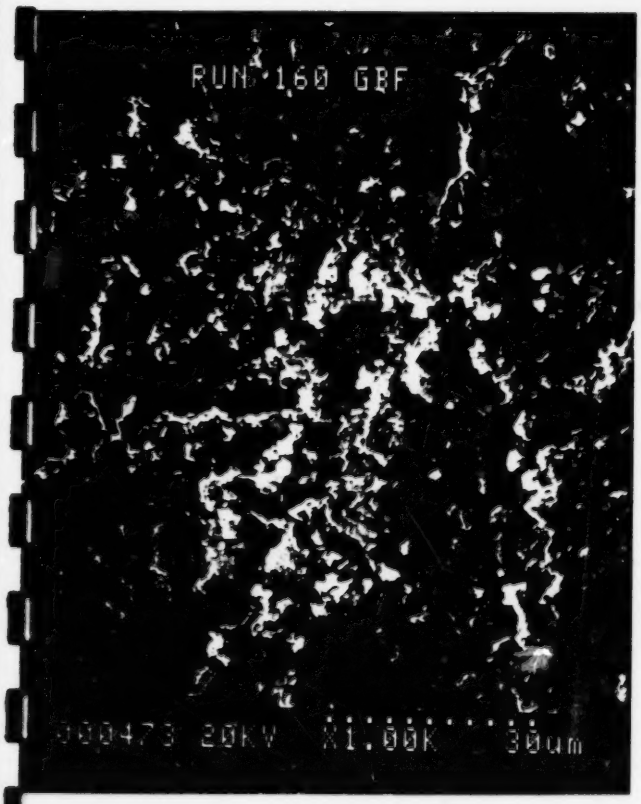
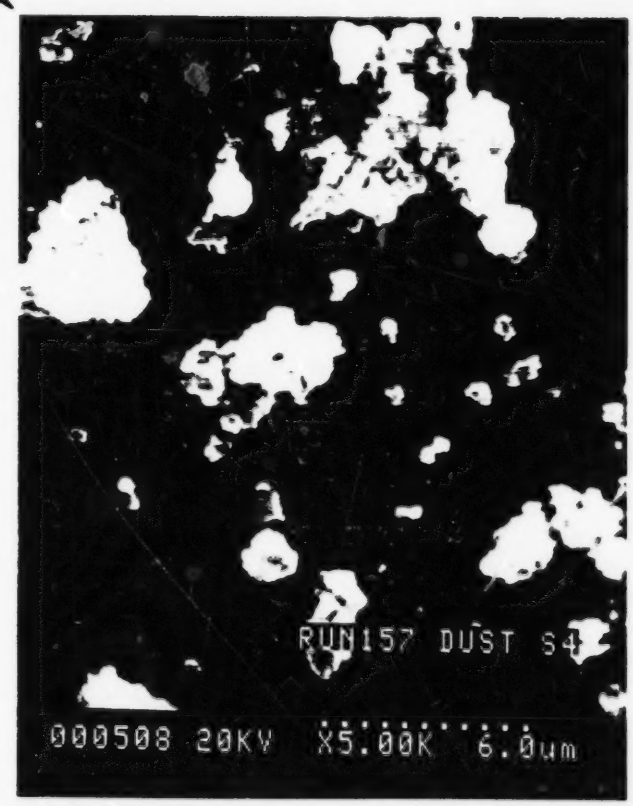


Figure 16 Particles Escaping Collection

(a) Run 144 x 2000
Anthracite - Std. Cycl. No. 1
followed by Vortex Pocket Cycl.

(b) Run 157 x 5000
Highvale - Vortex Pocket
followed by Gran. Bed

(c) Run 156 x 1500
Highvale - Vortex Pocket
followed by Gran. Bed



STATEMENT OF WORK

Background:

Commercial systems presently used to purify gasification products account for a significant portion of capital cost of gasification plants. These systems operate at low temperatures. The cooling of gasification products prior to their purification is associated with large heat losses. Efforts have been made to develop systems which could operate at temperatures about 900°C.

The design of high temperature cleaning systems requires a detailed knowledge of particulate and corrosive species types present in gasification products. Since Canadian gasification feedstocks, such as oil sand cokes, coal washery rejects and lignites have not yet been evaluated this kind of information is totally lacking. The main objective of this work would be to fill this gap.

You shall perform the following tasks:

Task 1 Sample Preparation

- obtain sample of Syncrude and Suncor coke, coal washery rejects from Sparwood plant and Onakawana lignite.
- perform the proximate and ultimate analysis
- dry and grind the samples to required mesh size

Task 2 Modification of Gasifier

The gasifier will be modified to allow the operation in a wide range of spouting velocities (up to 5 m/s). The modification will include:

- enlargement of the cyclone char recycle lines
- enlargement of oxygen and steam lines

Task 3 Design and Construction of Hot Gas Clean-Up Unit

- design and construct a vortex collector pocket cyclone, the design will consider a possible replacement of the primary cyclone as to ensure the operation with standard and the vortex cyclone in series;
- design and construct a granular bed filter, the design will consider a possible operation in a downflow static bed mode and a cross flow louvered moving bed mode.
- install a valved bypass line so that the secondary collection unit can be bypassed.

Task 4 Preliminary Testings

For this task use one high ash content and one low ash content feedstock, e.g., coal washery rejects and Suncor coke, respectively.

- perform a series of runs, i.e., five runs with each feedstock, with increasing spouting velocity up to the maximum one which can be handled.
- perform the following determinations: gas composition, gas heating value, gas yield and carbon conversion.
- evaluate the vortex collector pocket cyclone as a secondary and as a primary cyclone at velocities typical for normal and high throughputs.
- analyse the particles entering, captured and leaving the cyclone for size distribution using coulter counter and nanosizer techniques.

.../2

STATEMENT OF WORK (CONT'D)

Task 5 Evaluation of Granular Bed Filter

For this Task select an optimal primary cyclone (from Task 4). The granular bed filter assessment will include inert and active particles. The activities of this Task will involve:

- investigation of cross flow through beds of small (0.5 mm) versus large (2 mm) particles
- fixed versus moving bed
- active (half calcined dolomite) versus inactive (alumina) particles
- two different bed depths

For these evaluations the Suncor coke and coal washery rejects will be used as feedstocks.

Task 6 Evaluation of Feedstocks

On the basis of experimental results obtained in Tasks 4 and 5 an optimal set of experimental conditions and perform gasification runs with hot gas cleaning system in operation. Suncor coke, Syncrude coke, coal washery reject and Onakawana lignite will be used as the feedstocks. Two runs, one with a normal spouting velocity and the other with a high velocity, will be performed for each feedstock. During these tests the following determinations will be performed:

- gas composition
- gas heating value
- gas yield and carbon conversion
- size distribution of particles entering, captured and leaving the cyclone
- size distribution of particles entering, captured and leaving the granular bed filter
- carbon content of particles captured in cyclones and in the granular bed filter

About 100g sample of particles captured in cyclone and in the granular bed filter will be shipped to the Energy Research Laboratories in Ottawa.

Task 7 Report Preparation

The final report will give a detailed description of the work performed during this project. It should have the following format:

- Introduction
- Experimental
- Results and Discussions
- Conclusions and Recommendations

A technical presentation will be made by the contractor after the completion of this project. The exact date and place of this presentation will be decided by the CANMET scientific authority.

APPENDIX B

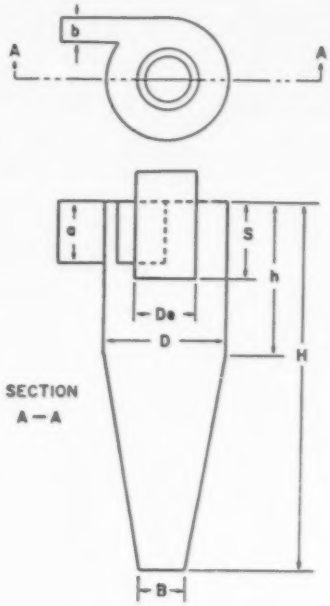
LEITH & LICHT EQUATIONS FOR CYCLONE EFFICIENCY

$$\eta = 1 - \exp\left(-2(C\psi)^{1/(2n+2)}\right) \qquad (1)$$

$$\psi = \frac{d^2 \rho_p v_i (n+1)}{18 \mu D} \qquad (2)$$

$$C = \frac{\pi D^2}{ab}\left(2\left(1 - \left(\frac{De}{D}\right)^2\right)\left(\frac{S}{D} - \frac{a}{2D}\right) + \frac{1}{3}\left(\frac{S+Z-h}{D}\right)\left(1 + \frac{d_c}{D} + \left(\frac{d_c}{D}\right)^2\right) + \frac{h}{D} - \left(\frac{De}{D}\right)^2 \frac{Z}{D} - \frac{S}{D}\right) \qquad (3)$$

$$d_c = D - (D - B)\left(\frac{S+Z-h}{H-h}\right) \qquad (4)$$



Dimensions of reverse-flow cyclone.

$$n = 1 - \left(1 - .67\, D^{0.14}\right)\left(T/283\right)^{0.3} \qquad (5)$$

$[D] = m$
$[T] = K$

Correction for Anisokinetic Sampling

A calculation is presented for a worst case, and a typical example. Badzioch's equation

$$W_s/W = (1-\alpha)V_s/V + \alpha$$

$$\alpha = [1 - \exp(-L/\lambda)]/[L/\lambda] \qquad L = .052 \text{ m}$$

$$\lambda = V_t V/g$$

Run 146 Wet gas composition: 6.1% H_2, 6.0% CO, 13.1% CO_2, 7.69% N_2, 0.12% CH_4 and 67.1% H_2O.

$T = 462°C$ $V_s = 105$ $V = 22.8$ m/s $(V_s/V = 4.605)$

(i) Calculate mixture viscosity and density by mol fraction average

$\mu_m = 2.699 \cdot 10^{-5}$ kg/m-s

$\rho_m = 0.3615$ kg/m^3

(ii) Assume $\rho_p = 1500$ kg/m^3 and calculate V_T and λ for various assumed particle sizes. Calculate α and W_s/W

d_p (μm)	V_T (m/s)	α	W_s/W	% Error in W
0.5	$7.57 \cdot 10^{-6}$	$3.385 \cdot 10^{-4}$	4.604	—
1.0	$3.03 \cdot 10^{-5}$	$1.354 \cdot 10^{-3}$	4.600	0.02
2.0	$1.21 \cdot 10^{-4}$	$5.416 \cdot 10^{-3}$	4.586	0.41
5.0	$7.56 \cdot 10^{-4}$	$3.334 \cdot 10^{-2}$	4.483	2.64
10.0	$3.03 \cdot 10^{-3}$	$1.353 \cdot 10^{-1}$	4.117	10.5
20.0	$1.21 \cdot 10^{-2}$	$4.561 \cdot 10^{-1}$	2.961	35.7

For particles 10 μm and smaller, the error in anisokinetic sampling can be ignored, for particles 20 μm, a 35% error is obtained. The measured concentrations are high and the efficiency of the cyclone will therefore be lower. conservative.

For a more typical case:

Run 147 Wet gas composition:

5.1% H_2, 5.7% CO, 11.1% CO_2, 7.5% N_2 and 70.5% H_2O

$T = 495\,°C$ $V_s = 55.8$ m/s $V = 25.0$ m/s $(V_s/V = 2.232)$

d_p (μm)	V_T (m/s)	α	W_s/W	% Error in W
0.5	$7.28 \cdot 10^{-6}$	$3.573 \cdot 10^{-4}$	2.232	—
1.0	$2.91 \cdot 10^{-5}$	$1.429 \cdot 10^{-3}$	2.230	0.09
2.0	$1.16 \cdot 10^{-4}$	$5.717 \cdot 10^{-3}$	2.225	0.31
5.0	$7.28 \cdot 10^{-4}$	$3.573 \cdot 10^{-2}$	2.188	1.97
10.0	$2.91 \cdot 10^{-3}$	$1.428 \cdot 10^{-1}$	2.056	7.9
20.0	$1.16 \cdot 10^{-2}$	$4.723 \cdot 10^{-1}$	1.650	26.
50	$7.28 \cdot 10^{-2}$	$8.722 \cdot 10^{-1}$	1.157	48

Here particles below 20 μm will give errors below 26%.

Appendix C Wet Gas Compositions and Solids Loading Calculation

WET GAS COMPOSITION

RUN NO.	WET GAS COMPOSITION (mol %)					
	H_2	CO	CO_2	CH_4	N_2	H_2O
135	14.04	7.75	26.83	0.20	11.71	39.47
136	21.93	13.52	21.48	0.04	9.20	33.83
137	23.11	11.98	29.55	1.10	13.46	20.79
138	19.37	11.76	19.36	0.56	8.60	40.25
139	11.25	9.66	13.84	0.45	0.14	64.67
140	12.08	10.81	13.82	0.32	0.27	62.62
141	8.98	9.73	17.51	0.20	6.69	56.76
142	6.62	7.24	14.60	0.13	9.77	61.95
143	6.34	5.96	11.55	0.10	7.42	68.95
144	4.32	3.57	12.27	0.12	3.39	76.31
145	9.24	9.00	13.40	0.16	6.97	61.22
146	6.06	6.02	13.08	0.12	7.69	67.06
147	5.10	5.67	11.08	0.14	7.50	70.50
149	9.51	10.17	14.32	0.17	5.25	60.58
150	6.53	9.58	11.82	0.16	50.54	21.38
151	0.00	0.00	9.96	0.0	72.90	7.86
152	3.18	3.54	12.24	0.10	51.41	29.51
153	1.52	1.23	12.04	0.03	53.43	31.77
154	8.16	9.48	12.02	0.53	54.60	15.21
155	4.92	6.50	10.74	0.27	46.63	30.94
156	7.30	11.33	11.96	0.38	56.81	12.21
157	6.28	10.00	11.31	0.34	51.84	20.23
158	15.33	10.97	15.60	0.20	8.12	49.78
159	5.93	8.32	13.65	0.16	61.00	10.95
160	4.46	5.25	12.03	0.00	50.79	27.39

SAMPLE CALCULATION - LOADING

Run No 151

Sample Point	S_1	S_2	S_3
Final Ceramic Filter Weight (g)	92.1169	67.6763	60.9161
Initial Ceramic " " "	52.0586	67.1518	60.3384
Wt. dust collected (g)	40.0583	0.5245	0.5777
Final Wet Test Meter Reading (ft^3)	189.945	156.745	234.228
Initial Wet Test Meter Reading (ft^3)	156.745	120.851	189.945
Vol. gas sampled (ft^3)	33.2	35.894	44.283
Time of sampling (min, sec)	46 : 41.38	34 : 29.10	36 : 13.63
Sampling Rate (ft^3/min)	0.711	1.041	1.222
Dust load (g / ft^3)	1.206	0.015	0.013
(g / n-m^3)	42.59	0.530	0.459

Run 151 gas flow 162 m^3 dry gas at 15°C / h
wet gas flow 198 kg / h

Gas loading at S_1 = $42.59 \frac{g}{m^3 \text{ dry } 15°C} \times 162 \frac{m^3 \text{ dry}}{h} / 198$ kg/h

= 34.8 g / kg = 34,800 ppmw

S_2 = 0.53 * 162/198 = 0.434 g/kg = 434 ppmw

S_3 = 0.459 * 162/198 = 0.376 g/kg = 376 ppmw

APPENDIX D

Recovery of Solids from Granular Bed Filter

Solids were recovered from the spent granular bed material as follows. A weighted quantity (about 250 g) of bed material was placed in a cylinder 7.5 cm dia. x 60 cm high partially filled with water containing 0.01 wt.% surfactant. Air was bubbled through a distribution for 20 minutes. The foam was collected and filtered through a 0.45 μm pore size Teflon filter, and the solids determined by weighing after drying.